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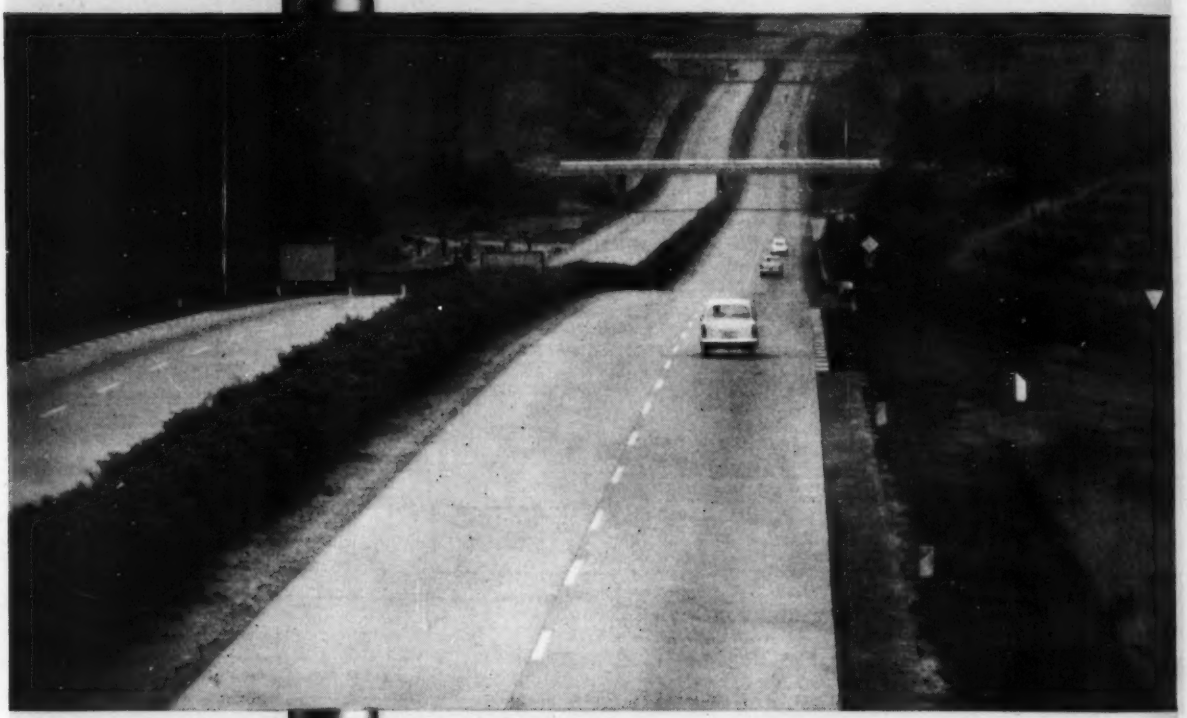

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APPLICATION OF HYDROREFINING IN THE PROCESSING OF VERTICAL RETORT TAR NAPHTHAS AND COKE OVEN CRUDE BENZOLES

By J. B. Lane

Technical Managing Director, Lancashire Tar Distillers, Limited

Communication No. 598

This Paper was presented and discussed at a Joint Meeting of The Institution of Gas Engineers, The Coke Oven Managers' Association, and the British Coke Research Association, at The Federation of British Industries, on 1st November, 1961. Mr. T. C. Battersby, M.B.E., M.I.Mech.E., M.I.Gas E. (President, The Institution of Gas Engineers), was in the Chair, supported by Dr. R. A. Mott, F.R.I.C., M.Inst.F., and Mr. F. E. Lodge, A.M.I.Gas E. (President and Honorary General Secretary respectively of The Coke Oven Managers' Association), Mr. G. W. Lee, M.Sc., and Mr. E. T. S. Wraith (Director and Secretary respectively of the British Coke Research Association; and Dr. A. E. Haffner B.Sc., M.I.Chem.E., A.R.C.S., D.I.C., and Mr. A. G. Higgins, M.Sc., F.R.I.C., (Vice-President and Secretary respectively of The Institution of Gas Engineers).

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I SUMMARY

Hydrorefining, both of vertical tar naphthas and coke oven benzole, is a fully developed process which can be turned into a reliable plant operation. It is no more difficult to operate hydrorefining process than acid washing. Perhaps, initially, the plant operation looks more complicated from an engineering aspect, but the chemistry and control is much easier because, once on stream, it is easy to keep the specification of products well in advance of those called for commercially.

There are no difficulties in producing pure benzene in high yields by distillation of a hydrorefined coke oven crude benzole.

The two main advantages of hydrorefining, against acid washing, are quality and high yield.

From similar crudes, the yield of water white chemicals and solvents is 20 to 30 per cent higher for tar naphthas, and 8 per cent higher for coke oven benzole. The quality of hydrorefined tar naphthas is vastly superior to that obtained by acid washing. Similar qualities of chemicals can be achieved by acid washing coke oven crude benzole, but only at much greater cost and sacrifice on yield.

There is no effluent problem with hydrorefining. Hydrorefining tar naphthas in cost per gallon of feed is more expensive than acid washing, but if related to the cost per gallon of water white solvents recovered it is about the same.

Prime costs of acid washing coke oven crude benzole to No. 2 pure benzole specifications are about the same or marginally lower than hydrorefining, but the capital cost and, thus, depreciation charge on hydrorefining is greater.

The overall realization of products less working costs for a tar naphtha by hydrorefining is 1·5 to 3·9*d.*/gal better than acid washing. The overall realization of products less working costs for a coke oven benzole by hydrorefining is 1·5*d.*/gal better than acid washing.

A pure benzole with a sulphur content less than 2 p.p.m., and perhaps less than 0·5 p.p.m., can be achieved in the future, both by hydrorefining and acid washing, although only by additional cost in the case of acid washing.

Pure benzole with crystallizing points in excess of 5·4°C can also be achieved both by acid washing and hydrorefining in conjunction with this low sulphur content. However, with such low thiophen contents the depression of the freezing-point will be caused by the aliphatic impurities such that the composition of the original crude benzole will be the most important function. Although the modest conversion of benzene to cyclohexane will accentuate this problem by hydrorefining, with a normal coke oven crude benzole at least 85 per cent of the benzene present in the crude can be recovered to this specification, the balance being to a lower crystallizing-point pure benzole and motor benzole.

However, if hydrorefining were supplemented by solvent extraction or "freeze" techniques, a minimum of 97 per cent of the benzene present in the crude could be recovered to a specification with a minimum crystallizing-point of 5·4°C.

Acid washing would also run into the same problem associated with aliphatic impurities of the crude benzole, and with the exception of crude benzoles with initial very low aliphatic content it is not possible to recover more than 83 per cent of the benzene present in the crude to a crystallizing-point of 5·4°C without a further supplementary process.

Solvent extraction developments of a hydrorefined vertical tar naphtha could yield high quality aromatics and paraffins at least equal to anything available in the world today.

Hydrorefining has a larger advantage the larger the throughput. The advantages at 10 mill. gal/a are greater than those quoted above, which are based on 6 mill. gal/a for tar naphtha and 4·8 mill. gal/a of coke oven crude benzole.

II INTRODUCTION

The Author first became actively interested in the hydrorefining process following a fire which completely destroyed the benzole refining plant at the Lancashire Tar Distillers, Limited, works at Cadishead in May, 1956.

At that time, the Coal Tar Research Association had proved the feasibility of the process in laboratory pilot plant, and the Newton Chambers pilot plant was under construction. However, at that time, there was no positive proof in the United Kingdom that the successful laboratory process could be successfully operated as a production unit.

A few years earlier, Benzole Producers, Limited, had produced reports regarding successful hydrorefining plants in Germany. These reports, without being conclusive, certainly stimulated the imagination to want to see more of them, and, accordingly, visits were made to Germany, where the Author was shown the plant at Scholven Chemie and a plant built by Lurgi at the factory of Matthias Stinners, and paid a visit to the Lurgi laboratories at Frankfurt, and a plant by Koppers at Minister Achenbach.

During the visit in October, 1956, the Author visited also the Dutch State Mines at Beek, and it was interesting to note that pulsed columns were being operated in the acid washing process, and that this technique was producing pure benzole to a specification below 1 p.p.m. The visit led to the conclusion that hydrorefining techniques could be successfully operated on plant scale, and that the laboratory figures could be successfully achieved, using both hydrogen and coke oven gas on coke oven benzole. However, the impression was gained that the chemical engineering and engineering problems were considerably greater than those of chemistry, and that the experience of Lurgi and Koppers in the design and construction of the plants had not been bought cheaply. In addition, the evidence showed that the process was thoroughly satisfactory for hydrorefining coke oven benzoles, but our problem was mainly on vertical retort tar naphthas. However, it was felt that, with the opportunity to start again, we had to try and resolve these difficulties and adopt a process that had such obvious advantage in quality and yield, even though the capital and working costs must work out higher.

The doubts we had about hydrorefining took a long time to dispel, but eventually, with the utmost help both from Lurgi and Koppers, the problem was resolved and an order was placed with Lurgi in May, 1958.

It was decided to use hydrogen rather than town gas, mainly because the pressure required to use town gas would have been of the order of 100 to 200 atm, as the local gas supply had a hydrogen content of only 30 to 35 per cent. We were fortunately aided in choosing hydrogen by the co-operation of Murgatroyd's Salt and Chemical Company, Limited, of Sandbach, who could supply cylinder hydrogen to our requirements at 99·9 per cent purity, and, although the price was not low, it did not necessitate additional capital expenditure on a hydrogen plant.

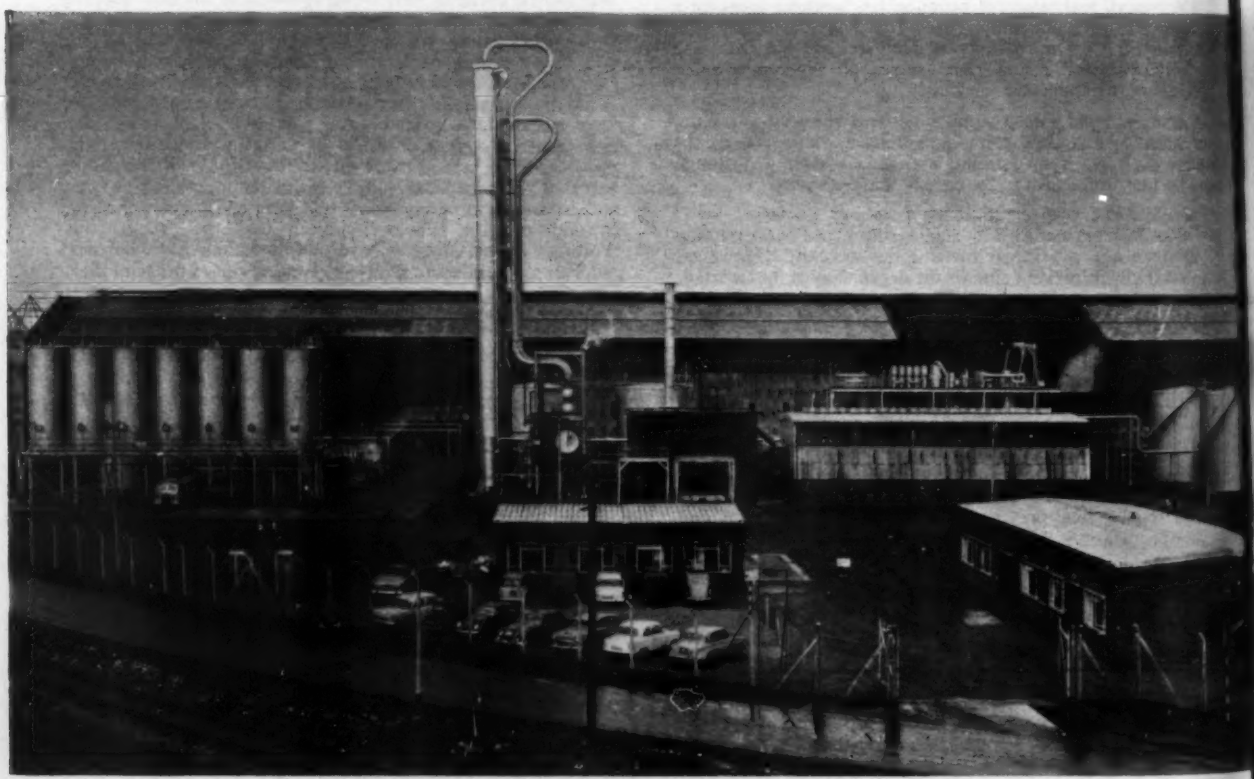
The plant was designed for 6 mill. gal/a, the throughput consisting of a mixed feed of 40 per cent crude gas works benzole and 60 per cent tar naphthas. The plant started up in June, 1959, and, thanks to Lurgi's starting-up team, we experienced few teething troubles. The only problem has

been to find sufficient raw material to keep the plant going due to the present trend of non-extraction of crude benzole at the gas works. The plant has been running at 60 per cent of its design throughput, and, to fill the plant capacity, we did buy some cracked petroleum feed, which gave considerable trouble in polymerization and coke deposits. The plant has done all that was expected of it, and the disappointing economic return has been caused by the low throughput. However, we have more than trebled our sales of refined tar naphthas since operating the plant.

In late 1958, the Lincolnshire Chemical Company, Limited, with which the Author is also associated, was asked to move the site of its works to make way for further expansion of the Lysaghts Scunthorpe Works (branch of G.K.N. Steel Company, Limited), whose crude benzole it refines. It was decided that while moving the old plant, which had been substantially modernized since the Second World War, the new plant would be brought completely up to date. However, there was in existence valuable modern acid washing plant and distillation plant, which could be re-used in the acid washing process.

Again the possibility of changing to hydrorefining was considered, but the facts of the case were different as at least half the existing comparatively new equipment would become redundant if hydrorefining were chosen, and the cost of modernizing by hydrorefining would be more than double the cost of modernizing by acid washing. The experience of hydrorefining at Cadishead was also not then to hand. Knowing that low-sulphur refined benzoles were produced by acid washing by the Dutch State Mines, and knowing that the A.P.V. Company was also actively pursuing similar developments with pulsed column washing, the decision was taken to modernize the acid washing to a 6 mill. gal/a plant constructed by the A.P.V. Company, using first an efficient defronting stage followed by a highly efficient continuous plant on the crude benzole to give a close-cut unwashed benzole and an unwashed residue from the crude benzole. The unwashed residue was further distilled to unwashed toluene, unwashed xylene and a 40 per cent indene fraction in a vacuum batch still. The unwashed fractions were then washed in batch washers, the washed benzole then requiring only a blow-over distillation, and the washed toluene and xylole a modest distillation up 36 plates. It was also known that, should pure benzole be required to parts per million of sulphur, a pulsed column could be used and was being developed, but there was no immediate hurry with existing specifications to complete this stage, particularly as the Lincolnshire Chemical Company management had other ideas that it wished to develop to supplement pulsed columns for producing low-sulphur benzoles. This step has never been regretted, and the Lincolnshire Chemical Company operations on the new plant, which started up in August, 1959, have been highly satisfactory.

In March, 1959, the Lincolnshire Chemical Company associated with the Steel Company of Wales, formed a joint company, the Port Talbot Chemical Company, Limited, to refine the crude benzole of the Steel Company of Wales, the plant design capacity being 4·8 mill. gal/a.



General View of the Port Talbot, South Wales, Works of the Port Talbot Chemical Company, Limited.

Here the plant would be a new construction on virgin ground, and hydrorefining against acid washing became a clear-cut choice. At that time, we tried to anticipate the future trend in pure benzole specifications, and, after consultation with Benzole Producers, Limited, decided to design the plant on a maximum yield of benzole with less than 5 p.p.m. sulphur, and a crystallizing-point of 5·35°C. There was some chance that a benzole would be required to this specification in the not too distant future.

The Cadishead plant was under way during these deliberations, and we were quite confident of the performance of the hydrorefining section. On the other hand, the Lincolnshire Chemical Company's new acid washing plant was working excellently, too. The decision depended, therefore, on the relative ease of distilling hydrorefined benzole, as there was a possibility that any cyclohexane produced in the hydrorefining process would cause difficulties in the production of pure benzene.

Here, again, the Lurgi Company was able to prove that this was no real problem, and showed us a combined hydrorefining/pure benzole distillation plant that it had built for Dortmunder Bergbau at Dortmund. The Lurgi Company supplied also a process design for the distillation unit, which was constructed and engineered by R. and J. Dempster, Limited, using Montz plates.

We were quite confident that in the hydrorefining section only a small amount of cyclohexane would be made, and that in the distillation section it could be removed easily in the forerunnings with the minimum amount of benzene. Orders were placed in October, 1959, and work began on site in January, 1960. The plant, with storage and all ancillary equipment, was finished in September, 1960, and went on stream on 2nd October, 1960. The speed of construction was considerably helped by Simon-Carves Limited, which did the plant erection both for Lurgi and Dempsters. The plant kept on stream, with the exception of shutdowns to train personnel, until Christmas, but was not opened for maintenance. It then ran through until the planned shutdown in July, 1961. During this running period, the desired yields and costs were achieved, with sulphur contents as thiophen averaging about 1·5 p.p.m.

As to the future, the real benefit of hydrorefining of vertical retort tar naphthas at Cadishead can be fully achieved only when we have developed a system of getting an aromatic paraffin separation for the whole hydrorefined product. Work is well forward on this project, but much has still to be proved.

The acid washing process at Lincolnshire Chemical Company, with its limit on yield, can be developed to give pure benzole below 0·5 p.p.m., at a little additional cost, by a process developed by the Company itself, and which is well advanced.

At Port Talbot, it is confidently expected that conditions of hydrorefining can be changed to give benzene with sulphur below 0·5 p.p.m. at a little extra cost, although, to give a crystallizing-point of 5·45°C, it may be necessary to recover the benzole from the forerunnings by solvent or "freeze" techniques, and would only be justified by a premium on this quality (or a discount on other qualities). However, the higher yield of products gives increasing advantage to hydrorefining over acid washing as the specification for pure benzene becomes tighter, provided the money can be found for the additional capital cost.

There is little need to elaborate further in the Introduction. This paper is a study of the working operations of two hydrorefining plants, their results and observations on the results, and a comparison with acid washing technique.

The first section deals with the operation of hydrorefining tar naphthas with pure hydrogen at Lancashire Tar Distillers, Limited, Cadishead, and the second section deals with hydrorefining coke oven benzole with coke oven gas at Port Talbot Chemical Company, Limited, at Port Talbot.

As to conclusions, each operator must draw his own, but the Author's conclusion is, that the higher the throughput, the tighter the specification of products and the less modern equipment in existence, the more positive becomes the appeal of hydrorefining.

The difficult decision is when the throughput is 6 mill. gal/a, or less: Does one modernize with hydrorefining, or cut one's expenditure and modernize with acid washing? Both processes can give the quality demanded for the future. Perhaps the correct trend is for large centralized refineries of 10 mill. gal/a, or greater, employing hydrorefining on the scale for which it was truly designed.

III HISTORICAL DEVELOPMENTS OF HYDROREFINING CRUDE BENZOLE AND NAPHTHA

(1) HYDROREFINING DEVELOPMENTS IN GERMANY

According to Nonnenmacher, Reitz and Schmidt from "Erdol und Kohle" 8th Year Book (1955), work on hydrorefining crude benzole was begun by the Badische Anilin and Soda Fabrik as long ago as 1925.

In B.A.S.F. work on the subject, it was stated that as long ago as 1925, in the pressure refining of crude benzoles, pressures of 10 to 100 atm, preferably in the region of 40 atm, could be employed, although it was customary to pressure-refine coal, petroleum and tar at 200 atm. In 1931, B.A.S.F., in a large-scale test pressure-refined 330 tons of crude benzole at 40 atm, using hydrogen, on a plant at Ludwigshafen. In 1933, a further large-scale test of 2,000 tons was tried out at Ludwigshafen, Thereafter, B.A.S.F. confined its activities to small investigations with different catalysts.

After the Second World War, Scholven Chemie A.G. began investigations on pressure-refining crude benzole and collaborated with B.A.S.F. In 1950, the Scholven Chemie A.G. at Gelsenkirchen set up the first large-pressure refining of crude benzole, using hydrogen and temperatures of about 350°C, and 60 atm pressure over a B.A.S.F catalyst. The plant has been progressively expanded to a throughput of 20,000 tons/month (1955).

In 1952, the Harpener Bergbau A.G. obtained a licence from B.A.S.F.-Scholven for the process, and a plant was erected at Amalia, near Dortmund, for 6,000 tons/month of crude benzole, using hydrogen.

The refining of crude benzole with coke oven gas instead of hydrogen was tried successfully by Scholven in 1951.

Co-operation began in 1951 with the Heinrich Kopper Company, which has since constructed a plant at Lothringen (under construction 1955). Co-operation began also with the Lurgi Company in 1952, in view of similar work it was doing in the petroleum field. In 1953, Lurgi erected a 500 tons/month plant at Nuremburg gas works, using town gas at about 40 atm pressure.

Since this reference in 1955, the B.A.S.F.-Scholven Process has been licenced to many operators throughout the world, and plants have been built by both Lurgi and Koppers, and many small developments perfected.

(2) COAL TAR RESEARCH ASSOCIATION WORK ON HYDROREFINING TAR NAPHTHAS

The Coal Tar Research Association began work, on behalf of its tar distilling members, in 1951 to develop a process for the complete hydrogenation of creosote fractions, with a view to obtaining a range of products from creosote suitable for high-boiling solvents, heat transfer medium, *etc.* Many catalysts were tested for the desulphurization of the tar oils prior to catalytic hydrogenation, but ultimately the fact emerged that cobalt molybdate catalyst was the best to be used. In general, it was found that, as the feedstock boiling-range increased, more drastic conditions were required to yield a sulphur-free and saturated product.

In order to obtain large enough quantities of hydrogenated creosotes for market research, a pilot plant consisting of a desulphurization stage and hydrogenation stage was constructed. This had a capacity of 0·5 gal/h and could operate at pressures up to 1,000 lb/in² gauge.

At this stage, they were unaware of developments in Germany other than reports in technical abstracts about developments by Scholven Chemie, where pressures of 150 atm were being used.

In 1953, at the suggestion of the Creosote Panel of the C.T.R.A., the hydrorefining process was tried on benzoles and tar naphthas, and desulphurization of these feedstocks was easier than with the higher-boiling tar oil and conditions either to give residual sulphur contents of a few parts per million, or to give products of good colour, colour stability and smell, but higher sulphur contents were established. One interesting point found in this development was that, to give the quality of refined product, a pressure of 400 to 500 lb/in² gauge had to be employed, and this led to a small degree of nuclear hydrogenation of aromatics. If however, the refined naphtha were washed with 65 per cent sulphuric acid before hydrorefining, or after hydrorefining, but before distillation, a milder hydrogenation treatment at 25 to 50 lb/in² gauge could be employed, giving equivalent quality of products. This development has been patented.

It came as something of a shock to the C.T.R.A. in 1954 to learn of the developments in Germany by B.A.S.F., and the plant designed and built by Lurgi at Nuremburg gas works, using a similar catalytic process.

Since then, Newton Chambers, Limited, being a member of the C.T.R.A., have developed the C.T.R.A. Process to 500 gal/d pilot plant, which commenced work in 1957 and has since operated successfully on a wide range of C.T.R.A. members' naphthas and benzoles ; no commercial plant has so far been built.

The C.T.R.A. has also made laboratory studies of taking a hydrorefined tar naphtha of vertical origin, (i.e., about 30 per cent paraffin content), passing it over an aromatizing catalyst which converts a substantial proportion of the paraffins to aromatics and yields sufficient hydrogen for the hydrorefining operation. This has obvious appeal as a further development at Cadishead.

Another approach was that the feed, before hydrorefining and mixed with steam, was passed over an iron oxide catalyst to yield hydrogen and carbon monoxide from the paraffins, this gas to be used for hydrorefining. This is also of some interest.

IV CHEMISTRY OF REFINING BENZOLE PRODUCTS

(1) CRUDE BENZOLE OR NAPHTHA

(A) AROMATICS

Under the mild conditions of hydrogenation using a molybdenum or cobalt-molybdenum catalyst (namely, pressures under 50 atm with a partial pressure of hydrogen about 18 to 20 atm and under 400°C), the aromatic benzene ring is hardly attacked at all, the conversions being :

Benzene ——→ *cyclo*hexane.
Toluene ——→ methyl *cyclo*hexane.

Naphthalene, when hydrogenated under these conditions, goes almost exclusively to tetralin, although at higher partial pressures it would lead towards further formation of decalin. Naphthalene is more readily hydrogenated under the same conditions than toluene, and toluene more readily than benzene.

(B) PARAFFINS AND OLEFINES

Olefines, di-olefines and cyclic olefines take up hydrogen to yield the saturated compound.

Paraffins remain unchanged as the temperature is too low for isomerization. Under the more severe conditions, it is possible for the side chains to be split from higher members of the *iso*-paraffins series.

Naphthenes can be dehydrogenated under certain conditions, especially with increasing temperature and decreasing pressure, but the equilibrium will normally be in favour of the hydrogenation of aromatics.

$$C_6H_{12} \rightleftharpoons C_6H_6 + 3H_2$$

(C) MISCELLANEOUS AROMATICS

Indene is hydrogenated to hydrindene.

Styrene is hydrogenated to ethyl benzene.

Coumarone is hydrogenated to ethyl benzene, but at very high temperatures the ethyl group can be split off as well.

Cyclopentadiene is reduced to cyclopentane

$$C_5H_6 + 2H_2 \longrightarrow C_5H_{10}$$

(D) TAR BASES

Pyridine bases are hydrogenated by being split at the ring, yielding paraffins and ammonia.

$$C_5H_5N + 5H_2 = C_5H_{12} + NH_3$$

Aniline yields ammonia and benzene.

Pyrrole yields butane and ammonia.

(E) PHENOLS

Phenols are reduced readily to hydrocarbon and water. At quite low temperatures, the hydroxy group is split off the multi-hydric phenols, to give the mono-hydric phenols. However, the mono-hydric phenol is not necessarily entirely converted to the corresponding aromatic under the conditions of our hydrogenation. Only above temperatures of 420°C is the equilibrium shifted substantially towards the formation of aromatics.

The hydroxy group activates the other carbon atoms to such an extent, particularly the neighbouring atoms, that the benzene ring is hydrogenated before the hydroxy group is split off, naphthenes being the main reaction product.

(F) REACTION OF SULPHUR PRODUCTS

Hydrogenation of sulphur products always leads to hydrogen sulphide. Elementary sulphur and carbon disulphide are very easily reacted ; this reaction needs a very low partial pressure of hydrogen and a low temperature.

Mercaptans are very much more stable, and higher temperatures are required for complete reaction especially if large amounts of hydrogen sulphide are present. Thiophens are the most stable under these conditions of hydrogenation. Thiophen (and its derivatives) are split at the ring, giving butane and hydrogen sulphide, the reactions going *via* tetrahydrothiophen and butyl mercaptan.

(2) GAS REACTION USING COKE OVEN GAS

(A) Type of Coke Oven Gas

When using coke oven gas as the hydrogenation medium at 50 atm over the molybdenum catalyst, the gas would be approximately 60 per cent hydrogen, other constituents being,

Carbon monoxide,
Carbon dioxide,
Unsaturated hydrocarbons,
Methane,
Nitrogen,
Oxygen,
Hydrogen sulphide.

There would be traces of other substances present. Methane and nitrogen do not react under the above conditions.

(B) Unsaturated Hydrocarbons

Unsaturated hydrocarbons dependent on the partial pressure of hydrogen are normally converted to the corresponding paraffin,

$$C_2H_4 + H_2 = C_2H_6 + 33{,}700 \text{ k. cal.}$$

(C) Carbon Monoxide

Carbon monoxide can react in different ways according to the conditions of the hydrogenation process.

$$CO + 3H_2 \rightleftharpoons CH_4 + H_2O + 50{,}420 \text{ k. cal.}$$

This reaction is pressure-dependent, and increased temperature favours methane formation ; alternatively, increased partial pressure of carbon monoxide favours methane formation.

As the above reaction yields water, the next reaction is

$$CO + H_2O \rightleftharpoons H_2 + CO_2 + 10{,}050 \text{ k.cal.}$$

or the two reactions combine,

$$2CO + 2H_2 \rightleftharpoons CO_2 + CH_4 + 60{,}470 \text{ k.cal.}$$

The carbon monoxide/water reaction is increased by the addition of water, which decreases methane formation especially when the pressure is below 50 atm.

Boudouard's reaction

$$2CO \rightleftharpoons C + CO_2 + 38{,}720 \text{ k.cal.}$$

is noticeable only with a high carbon monoxide content of the fresh gas. Increased temperature shifts the equilibrium in favour of carbon monoxide, but increased pressure has the opposite effect. This is a very undesirable reaction because of deposits on the catalyst.

(D) Carbon Dioxide

With the absence of water vapour at the catalyst, carbon dioxide can react to consume hydrogen :

$$CO_2 + H_2 \rightleftharpoons CO + H_2O - 10{,}050 \text{ k.cal.}$$

A high concentration of carbon dioxide in the fresh gas is therefore undesirable.

In the absence of carbon monoxide, carbon dioxide will react partly to form methane

$$\begin{aligned} CO_2 + H_2 &\rightleftharpoons CO + H_2O - 10{,}050 \text{ k.cal.} \\ CO + 3H_2 &\rightleftharpoons CH_4 + H_2O + 50{,}420 \text{ k.cal.} \\ \hline CO_2 + 4H_2 &\rightleftharpoons CH_4 + 2H_2O + 40{,}370 \text{ k.cal.} \end{aligned}$$

The resultant water is not sufficient to exclude methane formation altogether.

(E) Oxygen

Oxygen reacts to form water :

$$O_2 + 2H_2 = 2H_2O + 115{,}500 \text{ k.cal.}$$

(3) CATALYST

(A) Discussion

The Lurgi Process (being the engineering of the B.A.S.F. Process) employs the molybdenum catalyst on an alumina base. Other hydrorefining processes (notably the C.T.R.A./Newton Chambers/Esso research) employ a cobalt-molybdenum catalyst. Both seem effective.

It is the opinion of the Author that a cobalt-molybdenum catalyst gives a modestly better degree of de-sulphurization and could lead to slightly lower reaction temperatures necessary at the catalyst. It is claimed that a molybdenum catalyst is more effective in giving lower bromine numbers of the hydrorefined product, and the Author would not like to give an opinion on this.

The catalyst employed at both Cadishead and Port Talbot is an extruded one, which tends to be more reactive per unit volume than a pelleted one, and gives sufficient physical strength.

(B) Effects of Conditions on Catalyst Activity

With a cobalt-molybdenum operating in the range of 310° to 410°C, it shows that the efficiency of sulphur removal increases with temperature, but so also does the rate of catalyst deterioration when operating at atmospheric pressure.

Increase of the hydrogen partial pressure improves the desulphurization and retards carbon build-up on the catalyst. This favours using as high a hydrogen partial pressure in the plant as possible without seriously beginning the conversion of aromatics.

The gas ratio does not affect the desulphurization markedly except that slight improvements can be effected by increasing the gas : crude ratio at higher temperatures.

As might be expected, the lowering of the liquid feed rate (i.e., increasing the contact time) results in more complete removal of sulphur.

(4) EFFLUENT

One of the hidden advantages of hydrorefining is lack of effluent problem. This is not the case with acid washing. The effluent from the Cadishead plant is of the order of 70 to 100 gal/h, and, at Port Talbot, 120 to 150 gal/h of a water which has been in contact with both hydrogen sulphide and ammonia. The Cadishead effluent contains about 4 g/litre of ammonia and 3·3 g/litre of hydrogen sulphide. The Port Talbot effluent contains considerably less ammonia (due to low pyridine bases in crude), and has less hydrogen sulphide.

V HYDROREFINING TAR NAPHTHAS WITH PURE HYDROGEN

(1) DESCRIPTION OF PLANT

Crude naphtha is picked up at storage compound by pump (1) with enough pressure to pump through a fine filter (2) and then to the high-pressure pump (3) where it is compressed to the operating pressure, normally 28 to 32 atm. It is then passed through a heat exchanger (4) in liquid form against

the lowest heat from the hot gases from the reactor. The crude naphtha is preheated to about 100° to 110°C at this stage.

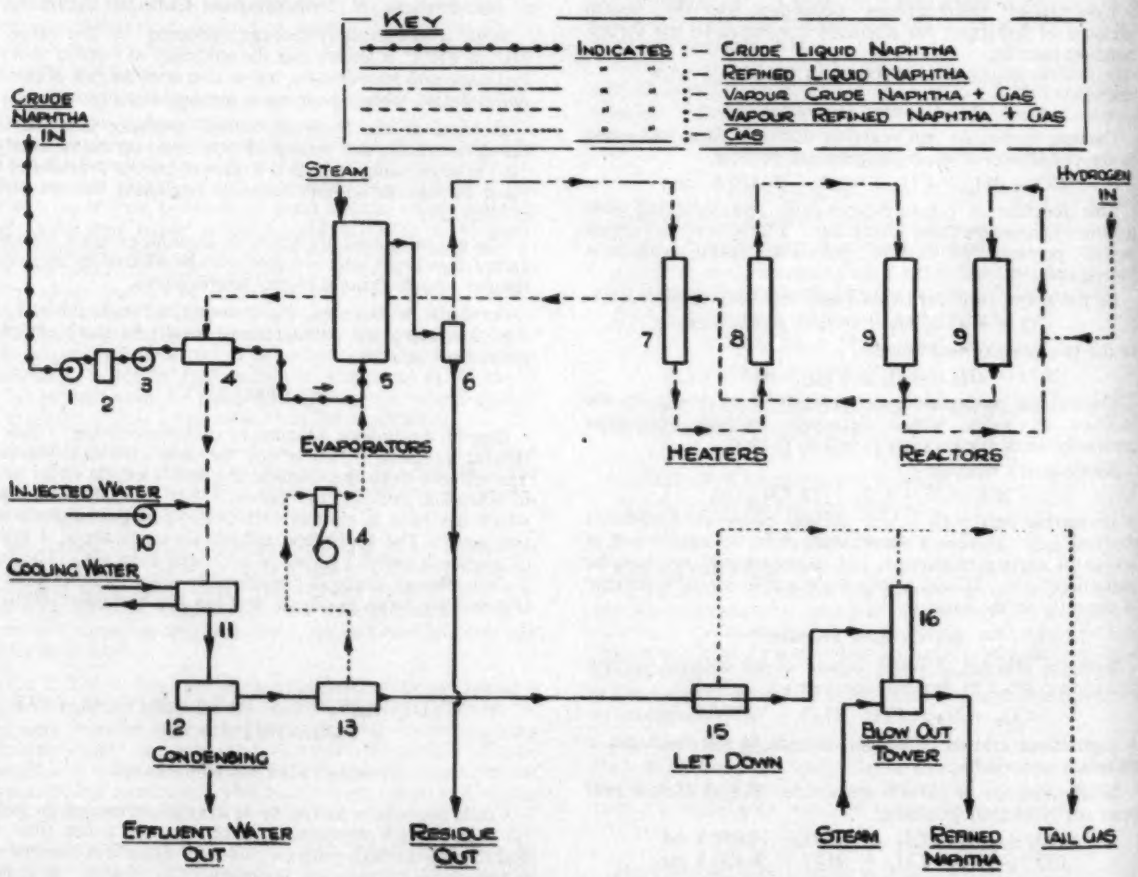
The liquid naphtha is then passed to a five-stage evaporator (5), where it joins the recycle gas, and the crude naphtha is evaporated into this recycle gas, leaving a residue of higher-boiling material. The heat for evaporation is supplied in four stages by heat exchange from the hot reactor gases, and in one stage by steam. The exit temperature in the evaporator, normally 190°C, controls the amount of evaporation. The exit temperature of the evaporator is used to control a by-pass arrangement of heat exchanger (7), by which more or less hot gases direct from the reactor are admitted to the evaporator stage.

The residue is run off through a small disengagement tank (6). The evaporated naphtha now in the recycle gas is then preheated to reaction temperature, first in a heat exchanger (7) with direct reactor gas, and secondly by a control electric heater (8), which is used for start-up mainly, and also when the total heat of reaction is not sufficient for preheat to catalytic temperature and evaporation.

The naphtha vapour recycle gas mixture is then introduced into the first stage of the reactor (9) at a minimum temperature of 300°C. The main part of the hydrogenation takes place in this stage with the hydrogen content of the recycle gas. The temperature of the gases rises 50° to 60°C in traversing the reactor bed. The hydrogen is introduced from high-pressure cylinders between the two reactors, through a pressure control valve set at the operating pressure, 28 to 30 atm, and the hydrogen flow is recorded. Hydrogen cools the reaction mixture by about 15°C. The hydrogenation is completed in the second reactor with a further gain of approximately 10°C, such that the final exit temperature is about 5° to 10°C lower than the exit from the first reactor, so that the overall temperature increase for the two reactors is about 45° to 55°C. However, these temperatures vary with constitution of the feed.

The hot reactor gases are then used to supply the heat to catalytic temperature in (7), then for evaporation in (5), and then preheat of the liquid crude naphtha in (4). The refined naphtha is then recovered from the recycle gas in the condensing system.

Firstly, cold water is injected into the gas stream by injection pump (10). The vapours are then condensed in a water-cooled condenser (11); the naphtha and injected water are then in liquid form, and the water is run off from the separator (12). The naptha and gas are then introduced in a seal pot (13) where the gas is separated and taken by recirculation compressor (14) back to the evaporators (5). The naphtha leaves the seal pot containing gases that are soluble at operating pressure, but will ultimately be released when the pressure is dropped to atmospheric.



Flow Diagram of Hydrorefining Plant for Treating Tar Naphtha with Hydrogen.

The refined naphtha then goes to the let-down vessel (15), where the pressure is dropped to about 5 atm, where most of the gaseous reaction products of hydrogenation are evolved as tail gas and are finally burned. The naphtha is boiled under reflux in a blow-out tower (16) operating at a few pounds above atmospheric, where the rest of the tail gas and hydrogen sulphide are removed. The pressure and temperature of the blow-out tower are adjusted so that the refined naphtha leaving the base of the column is free from hydrogen sulphide. The tower normally operates with a base temperature of 160°C.

A caustic and water wash system is incorporated to remove final traces of hydrogen sulphide from the refined naphtha, but as this is never used it is not shown on the flow diagram.

The hydrogen is purchased and transported in cylinders at 200 atm by road from Murgatroyd's Salt and Chemical Company, Limited, Sandbach. These are run down in process to about 40 atm and are returned for refilling. There is, therefore, no need for hydrogen storage or compressors at the plant. Each cylinder unit has a total capacity of 60,000 ft³ at N.T.P., so that each cylinder has an effective capacity of about 45,000 ft³ at N.T.P. because of the return pressure. This lasts about 4 h. There is always one cylinder unit in reserve on the plant, one on stream, one filling at Murgatroyd's, and one in transit, supplying about 240,000 ft³/d of hydrogen at N.T.P. to the plant.

(2) HYDROGEN CONSUMPTION

(A) Discussion

In view of the comparatively high cost of hydrogen, it is important to ensure the minimum consumption of hydrogen, and, therefore, to limit to the minimum any unnecessary loss in the process. Since there are limited day-to-day variations in constitution of the crude feed on stream, demanding varying amounts of process hydrogen, it is also important, for ease of control, that the hydrogen supply be controlled by a self-adjusting system and that there should not be a direct quantitative control relating hydrogen flow to hydrogen demand.

The throughput of the plant is governed by the catalyst contact time required for the various reactions to reach the desired degree of completion, and the availability of hydrogen. So long as the hydrogen supply and capacity of catalyst are not exceeded, reasonable variations in the constitution of the feed are automatically adjusted while on stream.

(B) Theoretical Hydrogen Requirements on Crude Analysis

To fix the hydrogen requirements, the following general figures for hydrogen consumption are used :—

	Hydrogen Uptake (ft³ at N.T.P./1,000 gal Crude)
0·1 per cent w/v sulphur as CS_2	225
0·1 per cent w/v sulphur as thiophen ..	450
1 g Br/100 ml	225
0·1 per cent w/v pyridine bases (as pyridine)	228
0·1 per cent w/v phenols	38
0·1 per cent w/v naphthalene to tetralin ..	56

The theoretical amount of hydrogen required can be calculated, and there are three potential losses :—

Hydrogenation of aromatics,
Hydrogen content of tail gas,
Plant losses through recirculation compressor, *etc.*

(C) Hydrogenation of Aromatics

To turn 1 lb of benzene to cyclohexane takes 14 ft³ of hydrogen at N.T.P. This is not only an expensive, but a physically undesirable, side effect. By fixing the operating pressures and temperatures as low as possible to achieve satisfactory quality of raffinate, hydrogenation of aromatics is minimized. It is difficult, due to the complex composition of the feed, to assess the amount of hydrogenation going on, particularly as there is cyclohexene in the crude, which is also hydrogenated to cyclohexane, but Table 1 shows an analysis by G.L.C. of the crude naphtha and hydrorefined naphtha over the process, showing a maximum conversion of benzene of 3·0 per cent, although this is not supported by the benzene balance, which would indicate the presence of cyclohexene.

Table 1.

Hydrorefined raffinate is 92 per cent of the feedstock.

		Feedstock	Total Hydrorefined Product
		Percentage w/w	Percentage w/w
1		2	3
Paraffins and olefins :—			
	C_5 and C_6	3·17	3·71
	C_7	2·65	2·97
	C_8	3·92	3·86
	C_9	4·91	8·41
	C_{10}	3·14	5·56
	C_{11}	3·75	3·71
	C_{12}	3·81	—
	cyclohexane	0·41	1·04
	methylcyclohexane	0·45	0·92
Aromatics :—			
	benzene	19·40	24·11
	toluene	13·80	14·50
	ethyl benzene	2·45	3·34
	m- and *p*-xylene	8·97	8·15
	o-xylene	3·90	3·06
	styrene	1·73	—
	n-propyl benzene	0·87	1·14
	ethyl toluenes	3·03	3·53
	mesitylene	2·37	2·52
	pseudo-cumene	3·52	3·06
	coumarone	3·46	—
	hydrindene	3·66	5·01
	indene	4·13	—
	tetralin	0·10	0·80
	naphthalene	1·60	0·60
Phenols		0·03	—
Carbon disulphide		0·20	—
Thiophen		0·57	—
		100·00	100·00

Another method of assessment is to account for the expected hydrogen consumption of the feedstock, and to compare the figure with that actually found on the plant. On the plant, the method used for the measurement of hydrogen is to take the pressure and temperature of cylinders before and after use and to calculate the volume discharged, converting to N.T.P., making allowance for the fact that hydrogen is not an ideal gas at the pressure involved. This method is considered sufficiently accurate for the commercial purchase of hydrogen.

On the guarantee run, the following hydrogen usage balance was obtained :—

Table 2.

	Amount In Crude	Hydrogen (ft.³/ton)	Percentage of Actual Usage
1	2	3	4
Sulphur (per cent w/w as CS_2) ..	0·41	206	5·9
.. (.. .. as C_4H_4S)	0·61	615	17·6
Bromine No. (g/100 g)	42	2,120	60·7
Basic nitrogen (p.p.m.)	233	9	0·3
Phenols (per cent, w/w)	0·3	22	0·6
Theoretical Use		2,972	85·1
Estimated H_2 content of tailgas ..		105	3·1
Naphthalene hydrogenated to tetralin (per cent, w/w).. ..	1·375	173	5·0
Loss, and monocyclics hydrogenated		240	6·8
Actual Use		3,490	100·0

If the total unaccounted-for hydrogen, 240 ft³/ton, were used for hydrogenation of aromatics (estimated as benzene) 17·3 lb/ton of benzene would have been converted to cyclohexane. This represents about 1 per cent of the aromatics present in the crude naphtha.

(3) PREPARATION OF THE FEED NAPHTHA

Table 2 also brings out the point of the necessity of careful preparation of the crude feed. The extraction of tar acids must be as complete as possible, not only because of their value, but because of their usage of hydrogen.

The case of tar bases is even more important. Although tar bases boiling up to 160°C have a ready sale and high commercial value, those boiling between 160° to 200°C have both a low market price and a limited market outlet, but it has been found that, in overall economics, it is cheaper to extract them and burn them than use hydrogen for their hydrogenation. As an illustration, hydrogenation of 2 : 6 lutidine requires 16·7 ft³ H_2/lb of 2 : 6 lutidine. The cost at 0·2*d.*/ft³ would be 3·34*d.*/lb, whereas extraction by traditional methods would be about 2·2*d.*/lb. The relative thermal values of heptane produced by hydrogenation and 2 : 6 lutidine are assumed to be approximately equal.

Close attention is also required to the naphthalene content of the feed. Fortunately, it is found that under our conditions naphthalene in the main is hydrogenated to tetralin, not to decalin, but, as can be seen, represent 5 per cent of our hydrogen usage. It is virtually impossible to eliminate the presence of naphthalene from the feedstock, but, by an efficient predistillation of tar naphthas, it is limited to about 2 to 3·5 per cent. Furthermore, naphthalene can be removed from the system in the residue from the evaporator (five-stage, no plates), but, whereas a sharp cut is not possible, by taking off a larger residue than necessary and redistilling off the naphtha content a degree of control can be achieved, getting about 25 per cent of naphthalene in the feed in the final residue.

Another trouble with the hydrogenation of naphthalene is that the heat of reaction is low compared to other impurity removal.

(4) METHOD OF REMOVAL OF GASEOUS PRODUCTS OF HYDROGENATION

The removal of the gaseous reaction products of hydrogenation relies on their solubility in the refined naphtha at the operating pressure of the plant in the seal pot which collects the liquid refined naphtha and releases the recycle gas to the recirculating compressor. The composition of the tail gas is determined by the relative solubilities of the various gases in the liquid hydrocarbon. On an evaporated throughput of 700 gal/h, the tail gas is about 2,500 ft³/h, but its quantity is, of course, dependent on the impurities in the feed. This compares to a hydrogen inlet of about 10,000 ft³/h.

A typical analysis of the tail gas, showing the major constituents, is as follows :—

	Per cent
Hydrogen sulphide	35 v/v
Hydrogen ..	15 v/v
C_nH_m	45 v/v

Taking the major constituents, their solubilities in hydrorefined naphtha at 30 atm and atmospheric are as follows :—

TABLE 3.

	Amount Dissolved at 30 Atm	Solubility at Atmospheric Pressure
1	2	3
C_nH_m	1,125 ft³/ft³	negligible
Hydrogen	375 ft³/h	negligible
Hydrogen sulphide ..	875 ft³/h	negligible

It will be seen that hydrogen sulphide is readily soluble, also being the product we wish to remove, whereas hydrogen, which it is not wished to remove at all, is the least soluble. Other light hydrocarbons are, naturally, the most soluble.

This loss of hydrogen via the tail gas, apart from leakage through recirculation compressor glands, *etc.*, is the only loss, the remainder of the gas being continually recycled from the seal pot. As described previously, the tail gas is released in two stages. Firstly, in pressure let-down from operating pressure to about 5 atm ; this is called expansion gas, and accounts for about 27 per cent of the tail gas.

The second release is from 5 atm to 10 lb/in² gauge in conjunction with the boiling of the naphtha in the blow-out tower to remove all traces of hydrogen sulphide from the naphtha raffinate. This is called blow-out gas, and accounts for the remainder of the tail gas.

Typical hydrogen sulphide contents of expansion, blow-out and tail gas are summarized as follows :—

TABLE 4.

	Tail Gas (Per cent, v/v)	Expansion Gas (Per cent, v/v)	Blow-out Gas (Per cent, v/v)
1	2	3	4
Make	100	27	73
Hydrogen sulphide content	30·3	8·4	38·3

(5) HEAT BALANCE

The heats of reaction of the two main reactions of hydrogenation (namely, conversion of thiophen and carbon disulphide to hydrogen sulphide and hydrogenation of the unsaturated compounds in the general range of tar naphthas) are sufficient to supply all the heat for the process and plant thermal losses.

This is on the assumption that the feed conforms to the following general type of specification :—

Total sulphur (w/v) (per cent)	= About 1
Sulphur as carbon disulphide (w/v) (per cent)	= About 0·4
Sulphur as thiophen (w/v) (per cent)	= about 0·6
Bromine No. (g/100 ml)	= 36 to 45
Percentage boiling at 100°C	= about 20

The total heat required for evaporation is somewhat greater the higher the boiling-range within the range 80° to 190°C, but, usually, so is the bromine number.

The following simplified estimate gives an idea of the amount of heat available from some of the reactions, based on calculated heats of reaction, and assuming the unsaturated molecules have only one unsaturated linkage.

It is assumed that the impurities of the feedstock are :

Sulphur as carbon disulphide ..	0·4 per cent (w/v)
Sulphur as thiophen	0·6 per cent (w/v)
Bromine No.	40 g/100 ml.

The total heat available from the formation of hydrogen disulphide, methane and butane from hydrogenation of

carbon disulphide and thiophen respectively, and the saturation of the double bond, calculated from bond energies would be about 180 Btu/lb of feedstock or about 1,500 Btu/gal.

Essentially, this is used to make up the radiation losses from the plant and the heat removed in the condenser. This heat has two main uses :

(*a*) to preheat the gases to reaction temperatures,
(*b*) to evaporate the feed.

These two functions are achieved as follows :—

The main heat exchanger, which raises the gas feed to reaction temperature by means of hot reactor gas, has a by-pass direct to the shell side of the evaporator, the evaporator, being normally heated by the relatively cooler effluent gas from the main heat exchanger.

However, a three-way valve, controlled by the temperature of the evaporator, opens or closes the by-pass valve so that the evaporator temperature is maintained at the desired level by addition of higher grade heat direct from the reactor by-passing the heat exchanger. In this way, the amount of evaporation from the feedstock can be controlled.

Extra heat in the form of an electric heater can be used to preheat the gases to the reactor, if the reaction heat is not sufficient to evaporate the required amount of feedstock and raise the vapour to reaction temperature. Steam, also, can be used instead of gas heat exchange on stages of the evaporator. A by-pass is also provided so that, if more heat of reaction is available than is required to evaporate and preheat the feed, some may be turned direct to the final water condenser.

It is most convenient to arrange the controls so that the gas feed to the reactor is fractionally below the desired temperature and very small amounts of electric heat, thermostatically controlled, is used to give the final boost.

The steam pass in the evaporator is more or less self-controlling, since it normally operates at the temperature of the steam, but under these conditions, with steam at only 160 lb/in^2, none, or very little, steam condenses. However, should the temperature of the pass fall, a temperature difference is established, steam condenses and the temperature is restored.

(6) LOSSES

(A) Discussion

The main advantage of the hydrorefining process is its high yields of products. Losses in any benzole or naphtha rectification process take place by two methods :

(*a*) physical loss of material shown by a material balance,
(*b*) side effects of the rectification process, leading to losses of the higher-priced material.

(B) Physical Loss by Weight Balance

The physical loss by hydrorefining is small, since the process is contained in a system which depends on an operational pressure of 28 atm, such that leaks, which might otherwise escape notice, cannot be tolerated from a safety aspect.

The losses involved in pressure let-down and hydrogen sulphide blow-out are also minimised by operating at the highest pressure and lowest temperatures, to keep the partial pressures of the recovered naphtha as low as possible, to keep the lowest concentration in the tail gas.

On the three-day guarantee run on measured quantities, the overall physical loss was as follows (no account taken of hydrogen weight) :—

	Per cent.
Loss (v/v)	1·08
Loss (w/w)	2·50
With sulphur as sulphur content w/v as carbon disulphide	0·39 per cent
With sulphur as sulphur content w/v as thiopen	0·61 per cent
Total sulphur (w/v)	1·00 per cent
Specific gravity	0·856

	Per cent (w/v)
The carbon disulphide impurity removed represents	0·54
The thiophen impurity removed represents	1·87
Leaving a weight loss by other sources	0·09

There is also a small drop in specific gravity of the hydrogenated products, which gives an apparent gain in volume over the process after allowance has been made for the removal of sulphur compounds.

(C) Losses due to Side Reaction Hydrorefining

The reactions taking place during hydrorefining are fairly clear cut.

The aromatics remain unchanged with the exception of some slight hydrogenation, and these hydrogenated aromatics (such as cyclohexane), although tricky to remove when pure benzole is the desired product, are recovered as motor spirit in the general boiling-range of the original feed.

The more important aromatic conversions are :

Indene (b.p. 182°C) ⟶ hydrindene (b.p. 177°C)
Styrene (b.p. 145·2°C) ⟶ ethyl benzene (b.p. 136°C)
Coumarone (b.p. 171°C) ⟶ ethyl benzene (b.p. 136°C)

The olefines are similarly saturated to the corresponding paraffins, which in general terms have somewhat lower boiling-points.

There is, therefore, a general slight lowering of the boiling-range of the feedstock.

(D) Losses Due to Side Reaction of Acid Washing

With the more conventional method of acid washing, one can expect the feed naphtha to have an increase in boiling-range after treatment.

The carbon disulphide is taken off as a " forerunnings ", usually with a larger quantity of light useful material ; in our case, we should expect this to be a minimum of 2·0 per cent of forerunnings, but in practice, because of the light nature of the feed naphtha, something nearer 4·0 per cent. These forerunnings would have little value.

Sulphonation of the aromatics would also take place, but this is most probably in line with the degree of hydrogenation, but, whereas the hydrogenation product has motor spirit value, the sulphonation product has none.

In acid washing, perhaps the greatest loss is due to the condensation and polymerization reaction in the presence of sulphuric acid. A degree of control can be exercized in the choice of acid quantity, acid strength, temperature of washing and time of washing, but there is no possible doubt that the loss is always greater than accounted for merely by removal of thiophen and unsaturated compounds. This loss appears not only as acid tar, but also as a high residue content in the distillation following acid washing. Before our hydrorefining plant came into operation, we tried to establish this figure, but a long series of tests gave no conclusive results. However, we believe it was between 10 to 30 per cent of the feed, and the subsequent yield of refined products, since the installation of our hydrorefining plant, leads us to believe it was of the order of 30 per cent. The condensation products were of a resinous nature and have a very limited outlet even as creosote.

(E) Recovery of Naphtha from Hydrorefining Plant Residues

The residue from the evaporator has only been separated by a flash evaporation and contains some naphtha.

Normally, the residue is run at a high percentage of the feed, say 8 per cent. This, to a limited degree, prevents naphthalene and other high-boiling material over 190°C being evaporated and hydrogenated, as these products could lead to coking in the plant, as well as using unnecessary hydrogen.

This residue contains about 75 per cent of naphtha boiling up to 190°C and is stripped in a stripping still, the recovered naphtha going back to crude naphtha feed and the remainder, representing about 2 per cent of the feed naphtha, being the true residue containing naphthalene, being returned to the tar plant where any remaining naphtha and naphthalene is recovered.

(7) YIELDS OF PRODUCTS FROM HYDROREFINING TAR NAPHTHAS

We can consider the yields from a typical feedstock of the following characteristics :—

Sulphur as CS_2 (w/v)	0·4 per cent.
Sulphur as thiophen (w/v)	0·6 per cent.
Bromine No. (g/100 ml)	40
Distillation-range :	20 per cent at 100°C ;
	50 per cent at 160°C ;
	92 per cent at 190°C.

At Cadishead, 20 per cent of the crude naphtha boiling up to 100°C is taken as a separate stream from the hydrorefining plant. The balance of the raffinate, being 70·5 per cent of the crude naphtha, has to be further distilled to sales grades naphthas leaving a distillation residue.

The hydrorefining plant residue contains a substantial naphtha content boiling up to 190°C. This naphtha is distilled off, leaving a true residue, the recovered naphtha going back to crude naphtha feed. The first yield from the hydrorefining plant as a percentage (v/v) of the crude naphtha feed is as follows :—

	Per cent.
Raffinate as motor benzole ..	20·0
Raffinate as naphtha for distillation	70·5
Residue for re-work (50 per cent up to 190°C)	8·0
Loss	1·5
	100·0

Thus, summarizing the yields from 100 gal of crude naphtha,

	Per cent.	Per cent.
(*a*) *Sales Products*		
Hydrorefining plant yield as motor benzole	20·0	
As raffinate feed for distillation ..	70·5	
		90·5
Less loss on handling and distillation	1·5	
Less distillation plant residues ..	4·0	5·5
Giving a yield of sales naphthas in one pass		85·0
Crude naphtha rework from Hydrorefining plant residue, 50 per cent of 8 per cent, recovering all of this in second pass		4·0
Ultimate overall yield of sales naphtha in sales grades after distillation		89·0
(*b*) *Residue*		
Residue from hydrorefining plant ..	8·0	
Less crude naphtha recovered by distillation of hydrorefining plant residues	4·0	
True residue	4·0	
Add raffinate distillation residue ..	4·0	8·0
(*c*) *Loss*		
Hydrorefining loss (v/v)	1·5	
Distillation loss (v/v)	1·5	3·0
		100·0

Comparing this with yields by acid washing, column 1 of Table 5 gives comparison for best yield by acid washing, and column 2 gives comparison for best quality by acid washing.

Table 5.

	Acid Washing		Hydrorefining
	Best Yield	Best Quality	
1	2	3	4
	Per cent	Per cent	Per cent
Yield of saleable naphthas ..	70	60	89
True residue ..	8	8	8
Polymerization residue	15	25	Nil
Loss by physical weight balance ..	7	7	3

(8) QUALITY OF REFINED PRODUCTS

In a refined tar naphtha, the quality is gauged more by degree of saturation of unsaturated compounds than the degree of desulphurization. Although the degree of desulphurization is not outstandingly important, hydrorefining does reduce the sulphur content of the raffinate to below 1 p.p.m. The degree of saturation is judged by four main characteristics of the refined product :

(i) bromine number,
(ii) acid wash test,
(iii) colour and colour stability,
(iv) favourable smell.

Here the distilled final product of hydrogenation has a firm advantage.

All the products have only their aromatic smell, and the colour stability is such that, even standing in clear glass bottles, they have remained water white after two years' exposure to light.

The bromine number and acid wash test of hydrorefined products from a tar naphtha is as follows :—

Table 6.

	Motor Benzole	Solvent Toluene	5° Xylol	Purosol 170	Purosol 190
1	2	3	4	5	6
Boiling-range ..	Up to 110°C	110°C	139° to 144°C	150° to 170°C	170° to 190°C
Acid wash less than	0·1	0·1	0·3	1·0	6·0
Bromine No. ..	0·07	0·08	0·56	about 1·0	about 1·0
Colour Lovibond yellow less than	0·1	0·1	0·1	0·1	0·1

With this type of naphtha, extracted from vertical retort tar, it is not possible by acid washing to approach this quality and even to get products that are comparable. This can only be achieved by substantial acid washing with the corresponding smaller yield of solvents and higher yields of distillation residue.

(9) AFTER TREATMENT OF PRODUCTS FROM HYDROREFINING PLANT

The inherent nature of vertical retort tar naphtha is its aliphatic content. In general, this is in the range of 20 to 30 per cent, and hydrorefining can do nothing to alter this fact. At present, the refined naphtha from the hydrorefining plant is split into sales grades by traditional straight fractionation.

A preliminary separation is made in a distillation column as part of the hydrorefining plant, into a motor benzole fraction boiling up to 100°C, and a naphtha fraction (boiling above 100°C). This latter fraction is distilled into sales fractions in a traditional vacuum batch still.

By using a column of about 45 theoretical plates and of low hold-up and an average reflux ratio of 5 : 1, it is possible to get distillation cuts where the aromatics in the desired boiling-range are concentrated with the minimum quantity of non-aromatics, and intermediate cuts of paraffin with the minimum concentration of aromatics, but the achievement falls far short of anything approaching a clear-cut separation.

It is obviously desirable, therefore, that some other method should be considered to achieve a further refinement of an aromatic stream and paraffinic stream. Many such methods have been considered, and all have been proved feasible within a limited range, and many within that range have shown economic advantage. However, none has yet been adopted, because of their limited application and because work in our own research laboratories led us to believe that, with reasonable luck, a process might yet be developed to treat the whole raffinate. The three main problems are :

(i) The high percentage of non-aromatics.

(ii) Most solvents preferentially dissolve the aromatics, which are still the main constituent.

(iii) That about half the material boils above xylene (150°C).

The third disadvantage is the most pressing as most world research and development is concentrated on raw material containing at least 85 per cent in the benzole—toluene—xylene range (i.e., boiling up to 160°C), thus, there has been no need for such a development. Secondly, the high-boiling nature of the feed causes difficulties due to the low boiling-point of the solvent, the instability at higher temperatures, or the low solubilities of the high-boiling aromatics in that solvent. This limits the scope of most of the known and developed solvent extraction processes, which deal with benzole, toluene and xylole only. Lurgi and Udex, to mention but two, have first class processes in this range.

The Newton Chambers "freeze" technique deals with the benzene fraction only. Here, again, it has been demonstrated that benzene of 5·4°C crystallizing-point can be manufactured easily and economically, but as this represents only 20 per cent of naphtha produced, it solves only the fringe of the problem.

Azeotropic distillation has been considered, but not only is the paraffin content too high for convenience, but again the problem of the high-boiling range is encountered.

The full advantage of superior quality cannot, therefore, be fully realized until the paraffin—aromatic separation can be achieved, but when this is achieved there is little doubt that both the aromatics and the paraffins will be at least equal in quality to anything being produced in the world today.

(10) ECONOMICS OF HYDROREFINING TAR NAPHTHAS

(A) Discussion

From the previous considerations, it will be seen that the main attraction of hydrorefining tar naphthas from vertical retort tar or benzole is in its yields and quality of products rather than in lower working costs.

(B) Running Costs of Hydrorefining Plant Using Hydrogen on Tar Naphthas

The running cost, based on 6 mill. gal/a of a typical tar naphtha of the following composition :—

Sulphur as carbon di-sulphide 0·4 per cent, w/v.
Sulphur as thiophen .. .. 0·6 per cent, w/v.
Bromine No. (g/100 ml) .. 40

with a distillation range of 20 per cent at 100°C, 50 per cent at 160°C, and 92 per cent at 190°C, would be as follows :—

TABLE 7.

Item	Unit/gal of Feed	Cost of Unit	Cost/gal of Crude (Pence)	£/a
1	2	3	4	5
Labour ..	1 man/shift	£3,600/year	0·144	3,600
Hydrogen ..	12·5 ft³	0·22*d.*/cu.ft.	2·750	68,750
Steam ..	3·0 lb/gal	100*d.*/1,000 lb.	0·300	7,500
Cooling water ..	24 gal/gal	4*d.*/1,000 g.	0·096	2,400
Condensate	5 gal/100 gal feed	2/6*d.* per 1,000 g.	0·002	50
Electricity ..	0·1 kW/gal feed	1·1*d.* per kW	0·110	2,750
Catalyst and Replacement ..	25 per cent/a		0·100	2,500
Maintenance	4 per cent of capital cost		0·400	10,000
		Prime Costs ..	3·902	97,550
Depreciation 10 per cent			1·600	40,000
Works overheads, namely, supervision, insurance, rent, rates, laboratory, yard, works general, say ..			0·400	10,000
		Overhead Costs	2·000	50,000
		Total Working Costs	5·902	147,550

Note: No provision for interest on capital.

(C) Summarized Costs for Tar Naphthas for Hydrorefining Distillation to Sales Grades and Residue Stripping

	Pence
1. Hydrorefining costs per gal	5·900
2. Assume cost of distillation of hydrorefined naphtha per gal of hydrorefined raffinate	2·000
3. Assume cost of residue stripping hydrorefining plant residues per gal of original residue feed	1·500
4. Assume cost of crude-unloading, storage, transfers and sales loading (excluding storage depreciation) per gal of crude naphtha	0·200

On these assumptions, the cost per gal to sales grades works out as follows :—

	Pence/gal
1. Hydrorefining plant costs	5·900
2. Distillation costs on naphtha raffinate, 70·5 per cent at 2*d.*/gal	1·410
3. Stripper costs on hydrorefining plant residues, 1·5*d.*/gal on 8 per cent of crude as residues	0·120
4. Hydrorefining plant costs on second pass on 4 per cent recovered crude naphtha from residue stripper, 4 per cent of 5·90*d.*	0·236

5. Distillation plant costs by second pass on 4 per cent recovered crude naphtha from residue stripper, 4 per cent of 1·410*d.*	0·056
6. Crude storage and receiving, pumping transfers, blending and sales loading..	0·200
Overall benzole plant costs per gal of original feed naphtha	7·922
Say	8·000
Yield of sales naphthas	89 per cent
Cost per gal of sales naphthas recovered..	8·99*d.*
Say	9·00*d.*

(D) Sales Realization, Less Costs, Comparing Hydrorefining with Acid Washing

Take, on average, sales values of naphthas, which includes motor benzole, toluene, xylene, solvent naphthas 90 to 160 and 90 to 190 at 30*d*/gal net, naked.

Taking the acid washing costs, including depreciation on a plant built at values of 10 years ago, it is estimated that the comparative working costs by acid washing would be 5·0*d.*/gal for best yield, and 5·5*d.*/gal for best quality.

Assume a value of true and polymerization residues at 9*d.*/gal net, naked. On these assumptions and on the yields mentioned previously, the relative economics would work out as follows, per 100 gal of feed naptha :—

TABLE 8.

	Acid Washing				Hydrorefining	
	Best Yield		Best Quality			
	Gal	Pence	Gal	Pence	Gal	Pence
1	2	3	4	5	6	7
Saleable naphthas at 30*d.*/gal ..	70	2,100	60	1,800	89	2,670
Residues at 9*d.*/gal	23	207	33	297	8	72
Loss	7	—	7		3	
Revenue from sales	100	2,307	100	2,097	100	2,742
Less working costs	100	500	100	550	100	800
Net surplus ..		1,807		1,547		1,942
Advantage for Hydrorefining (*d.*/100 gal) ..		135		395		—

It is also interesting to note that, although the working cost per gallon of crude naphtha, compared to acid washing, is higher, the cost per gallon of sales naphtha produced works out as follows :—

TABLE 9.

	Acid Washing		Hydro-refining
	Best Yield	Best Quality	
Yield of sales naphtha ..	70 per cent	60 per cent	89 per cent
Working cost/gal naphtha	5·00*d.*	5·50*d.*	8·00*d.*
Working cost/gal sales naphtha	7·1*d.*	9·2*d.*	9·00*d.*

(E) Conclusion

The advantage from hydrorefining is, therefore, not only in the superior quality, but also in that higher yields of saleable naphtha more than balance the higher working costs. The advantage financially lies between 1·35*d.*/gal to 3·95*d.*/gal for an altogether superior product, which, on 6 mill. gal/a represents £33,750 to £98,750/a. Assuming the distillation and storage is in existance and only the hydrorefining plant needs erecting at today's costs, an expenditure of approximately £300,000 is needed, giving a return on capital of 11 to 33 per cent. This ignores any premium for the superior quality, which might take the form of a monetary premium or a physical premium of selling products of a superior quality at the same price, where terms of trade are weak.

VI HYDROREFINING COKE OVEN CRUDE BENZOLE WITH COKE OVEN GAS

(1) DESCRIPTION OF PLANT

The crude benzole is picked up at the tank farm by pump (1) and passed through a filter (2), and then picked up by high-pressure pump (3) where it is delivered into the plant at the operating pressure of 40 to 50 atm. It is met by the recycle gas and passed to the six-stage evaporator (4) where the crude benzole is evaporated into the recycle gas, using the hot gases from the reactor in heat exchange as the heating medium together with one stage on steam. The evaporation is controlled at 190°C by by-passing part of the hot reactor gases around the next heat exchanger (7). The liquid residue is withdrawn through vessel (5).

The crude benzole vapour recycle gas mixture goes forward through a pre-catalyst (6) and through heat exchanger (7) against the hot gases from the reactor, then forward to the catalytic heater (8). The catalytic heater derives its heat from the catalytic non-flame burning of the tail gas, the temperature being controlled by the quantity of air added by air compressor (9) to the stream of tail gas leaving the plant. This heated tail gas is used in heat exchange against the crude benzole vapour recycle gas mixture to bring it to reaction temperature. The whole is built as one unit in six stages. The crude benzole vapour recycle gas mixture is thus heated to 340° to 350°C where it enters the reactor (10). The reaction takes place in the top part of the reactor with the recycle gas, the temperature increasing to about 380°C. At the half-way stage, fresh coke oven gas is introduced through a scrubber (11) to a four-stage compressor (12). Between the second and third stages of the compressor the coke oven gas is passed through a pre-treatment section (13), to remove traces of oxygen and other undesirable constituents. The coke oven gas is then compressed in the last two stages to the operating pressure of the plant and introduced into the reactor at 150°C. The reaction is completed in the lower half of the reactor, leaving at a temperature of about 370° to 385°C. The crude benzole is now refined and the hot reactor gases (being refined benzole in coke oven gas) pass through heat exchanger (7) and evaporator (4). Water is then injected into the stream of refined benzole vapour/coke oven gas mixture by injection pump (14) and is cooled in a water cooler (15). The refined benzole is then in liquid form, and it passes to an expansion vessel (16), where the recycle gas is removed. Part of this gas is picked up by recirculation compressor (17) and re-introduced into the evaporator ; part, balancing the coke oven gas input, is withdrawn from the system through a water cooler (18) and then let down in pressure by a valve (19) to about 80 lb/in² gauge. It then by-passes electric heater (20) (used for starting up) and goes through heat exchanger (21) to catalytic heater (8), where it is burned catalytically as described before. After this, the hot tail gas goes through heat exchanger (21), preheating the incoming tail gas to catalytic heater, through a water cooler (22) through a let-down valve to 15 lb/in² gauge (23) to the final gas line.

The liquid refined benzole leaves let-down vessel (16) and goes to the water separator (24), where effluent water is run off and pressure reduced to 150 lb/in² gauge, releasing some gas. The pressure is further let down to 70 lb/in² gauge in seal pot (25) and introduced to hydrogen sulphide blow-out tower (26), where the hydrogen sulphide is completely removed, and joins the final gas. The refined benzole is withdrawn as a bottom product from the blow-out tower, the expansion gas from (24) (25) and the blow-out gas (26) being let down to 15 lb/in² gauge by valve (27) before joining the final gas stream.

(2) YIELD OF PRODUCTS FROM HIGH-GRAVITY CRUDE BENZOLE

Hydrorefining of coke oven benzole with coke oven gas gives a very good yield of refined products over the process. The losses are in two parts,

(i) physical loss by weight balance,
(ii) loss due to side reactions in rectifying process.

(A) Physical Loss by Weight Balance

The physical loss is greater than when using hydrogen because the final gas leaving the plant at approximately 15,000 s.ft³/h (or approximately 24 s.ft³/gal of crude benzole refined) is saturated with benzene, which is not worth recovering on such a small scale.

However, on observed figures during the guarantee run, the loss, w/w, was below the guarantee 2·5 per cent, w/w, averaging about 2·0 per cent of the crude feed. The loss, v/v, was of the same order.

Regular plant-running figures, v/v, indicate that the physical loss, v/v, of the plant is about 2·0 per cent, v/v, or lower.

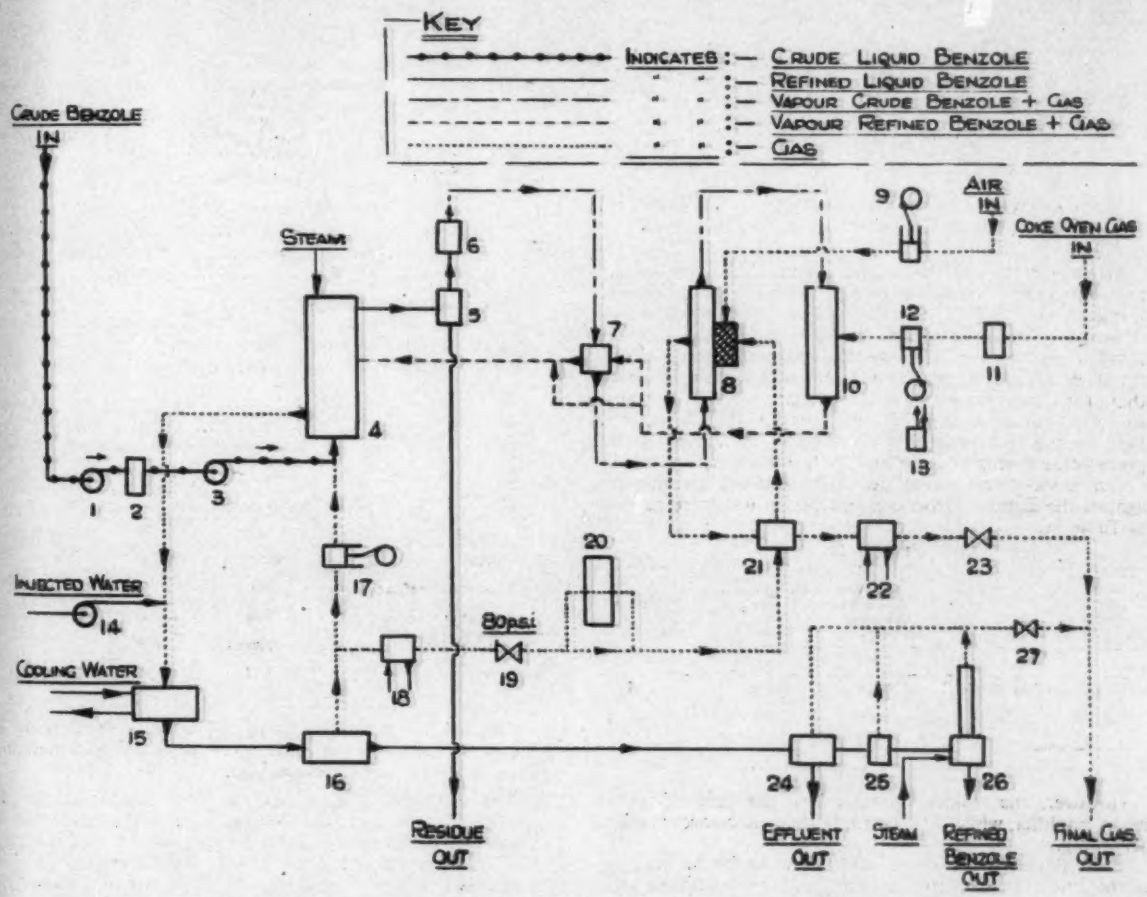
This loss includes the loss of sulphur impurities which, in the case of this crude, are as follows :—

Per cent (w/v)	Per cent (w/w)
0·21 of sulphur as carbon disulphide represents	0·29
0·23 of sulphur as thiophen represents	0·71
Physical loss due to removal of sulphur products	1·00

The loss also includes the loss of cyclopentadiene from the crude, which becomes a gaseous product on hydrogenation.

(B) Loss Due to Side Reaction

The loss due to side reaction is very low. Polymerization can take place only in the evaporator, and this goes on to a very limited degree as the plant can go without cleaning in this section for periods greater than six months. The side



Flow Diagram of Hydrorefining Plant for Treating Crude Benzole with Coke Oven Gas.

reaction of hydrogenation of benzene to cyclohexane, and toluene to methyl cyclohexane, does go on to a very limited degree. However, these hydrogenated products, although a disadvantage in the recovery of pure benzene, do at least have motor spirit value.

The hydrogenation of benzene represents under 0·3 per cent, v/v, of the benzene present, and toluene under 1 per cent, v/v, of the toluene present under our conditions of hydrogenation.

Table 10 gives the G.L.C. analysis of the crude benzole against yield of products from the plant, being hydrorefined raffinate and residue.

TABLE 10.

Raffinate equals 95 per cent of crude benzole.

	Crude Benzole (Per cent, w/w)	Total Raffinate (Per cent, w/w)
1	2	3
Paraffins and olefins including *cyclo*pentane, *n*-hexane, *n*-heptane ..	0·45	0·87
*cyclo*hexane	0·11	0·30
Methyl*cyclo*hexane	0·02	0·15
Benzene	61·42	64·63
Toluene	21·35	18·79
Ethyl benzene	0·30	1·91
m- and *p*-xylenes	7·44	6·77
o-xylene	1·33	1·41
styrene	1·33	—
ethyl toluenes	0·21	0·35
mesitylene	0·55	0·68
pseudo-cumene	0·71	0·68
hydrindene	—	1·00
indene	1·14	—
tetralin	—	1·34
naphthalene	3·20	1·10
Carbon disulphide	0·21	—
thiophen	0·23	—
	100·0	100·0

(c) Comparison of Hydrorefining and Acid Washing Yields Against Benzole Producers Evaluation

When looking at the overall yield from hydrorefining and subsequent distillation to pure benzene, motor benzene, nitration and "90's", toluene, xylene, solvents and residue it would be well to compare the yield of products against a familiar test, namely, the Benzole Producers evaluation test for crude benzole to motor benzole. It must be remembered that this defines motor benzole as that material boiling up to 160°C (to be specific, 95 per cent, at 155°C) where, in fact, in practice both in hydrorefining and acid washing processes, solvents or naphthas are recovered up to 190°C.

Over a six-month period, the works yields on hydrorefining against the Benzole Producers evaluation test were as given in Table 11.

TABLE 11.

	Benzole Producers Evaluation (Per cent, v/v)	Hydrorefining Yields (Per cent, v/v)	Variation (Per cent)
1	2	3	4
Products up to xylole 160°C	88·19	89·85	+1·66
Residue	7·88	6·79	−1·09
Loss allowance	3·93	3·36	−0·57

However, the residue includes 2·60 per cent of refined heavy naphtha, which would reduce the true creosote residue accordingly.

These figures are achieved after allowing for the recovery of the crude spirit content to 190°C from hydrorefining plant residues shown in Table 12. Although the hydrorefining plant residues are only 4 per cent, v/v, of the crude, they contain 75 per cent, v/v, as crude spirit to 190°C, which has to be recovered and recirculated to the crude benzole feed. Thus, in total, the hydrorefining residue shown in Table 11 is accounted for as follows :—

	Per cent.
Recovered as rectified heavy naphtha ..	2·60
Balance as creosote residue	4·19
	6·79

The total residues are accounted for in Table 12 :—

TABLE 12.

	Hydrorefining Plant Residues (Per cent, v/v)	Distillation Plant Residues (Per cent, v/v)	Total (Per cent, v/v)
1	2	3	4
First production (per cent, v/v) crude	4·09	5·79	9·88
Accounted for as follows:—			
Recovered as crude benzole (per cent, v/v) of crude	3·09	—	3·09
Recovered as high-boiling solvents (per cent, v/v) of crude	—	2·60	2·60
True Residue (per cent, v/v) of crude ..	1·00	3·19	4·19

On a modern acid washing plant making similar-grade products, including pure benzole to No. 2 specification, one would expect the following yields :—

	Per cent.
Products to xylole (160°C)	81·7
Residue	10·8
Loss	6·0
Carbon disulphide forerunning	1·5

Again, the residue would include 2·6 per cent of a refined heavy naphtha part recovered as an unsaturated naphtha fraction, making the true creosote 7·2 per cent.

The comparative yields, making pure benzole, assuming identical crude benzole by hydrorefining and acid washing and conforming to motor spirit allowance in Benzole Producers evaluation formula, are as given in Table 13.

TABLE 13.

	To Motor Spirit (B.P. Evaluation) (per cent, v/v)	To Pure Chemicals	
		Hydro-Refining (per cent, v/v)	Acid Washing (per cent, v/v)
1	2	3	4
Products up to xylole ..	88·19	89·85	81·7
Residue as rectified naphtha	Nil	2·60	2·6
Residue as creosote ..	7·88	4·19	8·2
Forerunnings	Not applicable	Nil	1·5
Loss	3·93	3·36	6·0

The loss of products up to xylole by acid washing is, therefore, about 8 per cent more than by hydrorefining, and could bear some examination.

The yield on hydrorefining is most spectacular in the xylole range, where about 5·0 per cent of the crude can be recovered, compared to about 3·0 per cent by acid washing, accounting for about 2 per cent of the difference. This is particularly severe in acid washing in crudes of high styrene content, where a styrene xylole complex gives losses of both styrene and xylole.

The light paraffins and benzene in the carbon disulphide forerunnings of an acid washing process account for a further 1 per cent.

The additional yield, therefore, by hydrorefining in the benzene and toluene cut is about 5 per cent of the crude and is accounted for, firstly, by losses of benzene and toluene, both by sulphonation, where the product goes to the residue, and secondly by polymerization of unsaturateds and thiophen, which, in a complex polymerization effect, additionally account for both benzene and toluene in these compounds, and thirdly by the overall greater physical loss by weight balance in the process.

(3) QUALITY OF REFINED PRODUCTS

Hydrorefining does give a high quality of refined products. The desulphurization is completed to give a sulphur content of under 2 p.p.m., which is well ahead of present British commercial standards. However, recent developments on acid washing process have made it possible to achieve even more complete desulphurization of the order of 0·5 p.p.m. or less at a slightly greater cost and lower yield, and so hydrorefining is not the only method of achieving low-sulphur benzene, but it is, of course, easy, cheap, reliable and a high-yield method. Only the future can tell what ultimate degree of desulphurization might be required. The present British standard of 100 p.p.m. as thiophen, or the German of 10 p.p.m. as sulphur, must be looked upon as interim specifications until the market settles down to petroleum competition with standards of 1 p.p.m. or less. Colour stability, odour and acid wash test, are in general better than by acid washing, but little ahead on modern acid washing methods.

(4) AFTER-TREATMENT OF HYDROREFINED COKE OVEN CRUDE BENZOLE

(A) Discussion

After hydrorefining crude benzole, it is necessary to distil the hydrorefined raffinate to the appropriate pure chemicals, namely, pure benzene, pure toluene, "90" toluene, 3/5° xylene, and higher boiling solvents.

As between 65 and 70 per cent of the hydrorefined raffinate is benzene, the main problem is concerned with this product. One of the features of the hydrorefined raffinate, unlike an acid-washed one, is that the apparatus can be constructed in mild steel, with no risk of corrosion, and this enables substantial construction on distillation column plates at very reasonable capital cost.

The hydrorefined raffinate contains the paraffins present in the crude benzole (which might go in the carbon disulphide forerunnings in an acid washing system) together with any cyclohexane and methyl cyclohexane converted from benzene and toluene by hydrogenation. The distillation is, therefore, not just one of light defronting followed by a straightforward benzene—toluene split as in acid washing, but needs a considerable defronting of the benzene, particularly as the cyclohexane azeotrope with benzene boils very close to that of pure benzene.

(B) Method Used at Port Talbot

The method of after-treatment adopted at Port Talbot is by efficient defronting by continuous distillation. Montz plates are used throughout. The main section of the column consists of 50 plates, where the main benzene—toluene split is effected, the toluene plus material leaving the base of the column, while on top of this column is a defronting section consisting of 25 plates from which forerunnings are taken overhead. The benzene offtake from the top of the main column is led to a side-stripper consisting of 60 plates, the top product of this returning to the defronting section, while pure benzene is taken from the base of the side-stripper. Only one reflux is used for both the defronter and benzene—toluene split. The toluene plus material is redistilled in a batch still, but if the throughput were greater this would, of course, be distilled continuously for toluene, xylene and solvents.

The quantity of forerunnings produced is dependent on the quality of pure benzene required. Making pure benzene to No. 2 specification of minimum 5·2°C crystallizing-point (normally made at 5·27°C), the quantity of forerunnings is about 4 per cent of the hydrorefined raffinate. The plant has made benzole at 5·45°C crystallizing-point where the forerunnings rises to about 15 per cent of the hydrorefined raffinate.

(C) Steam Consumption

The steam consumption per gallon of hydrorefined raffinate is also dependent on the purity of pure benzene, being 7 to 8 lb of steam/gal of hydrorefined raffinate in the benzene recovery section when making No. 2 specification, and 10 to 12 lb of steam/gal of hydrorefined raffinate when making 5·45°C crystallizing-point.

In addition, the toluene plus (about one-third of the hydrorefined raffinate) has to be distilled in a vacuum batch still where the steam consumption is between 9 to 15 lb of steam/gal of toluene plus (dependent on the quality of products required) equivalent to 3 to 5 lb of steam/gal of hydrorefined raffinate. This makes a total steam consumption of 12 to 20 lb/gal of hydrorefined raffinate or 11 to 18 lb/gal of crude benzole.

A typical G.L.C. analysis of the products from the continuous benzole plant is given in Table 14.

Table 14.

	Forerunnings	Pure Benzene (c.p. 5·33°C)	Toluene Plus
1	2	3	4
Percentage (v/v)	4·0	62·0	34·0
	Per cent	Per cent	Per cent
iso C_6 paraffins	0·14	Nil	Nil
n C_6 paraffins	0·43	Nil	Nil
iso C_6 paraffins	0·78	0·050	Nil
n C_6 paraffins	12·92	Nil	Nil
iso C_7 paraffins	4·17	Nil	Nil
n C_7 (heptane)	0·12	0·113	0·1
*cyclo*hexane	5·50	0·144	Nil
Methyl*cyclo*hexane ..	Nil	0·086	1·0
Benzene	74·34	99·592	7·6
Toluene	Nil	0·015	45·3
Ethyl benzene	Nil	Nil	4·5
m.p. xylene	Nil	Nil	15·4
o xylene	Nil	Nil	2·6
C_9 paraffins	Nil	Nil	0·6
C_9 paraffins	Nil	Nil	0·3
C_9 aromatics	Nil	Nil	2·2
Mesitylene	Nil	Nil	1·4
Hydrindene	Nil	Nil	7·8
Tetralin	Nil	Nil	4·1
Naphthalene	Nil	Nil	5·3

(D) Discussion on Table 14

From this, it can be seen that the benzene content of the forerunnings represents about 3 per cent of the hydrorefined raffinate, or just under 5 per cent of the benzene, present in the feed. The benzene content of the toluene-plus fraction is recycled to the feed of the main benzole column. The forerunnings have a specific gravity of about 0·834.

It is apparent that the defronting section is very effective in dealing with the cyclohexane, almost three times as much going to the forerunnings as to the pure benzene. The *n*-heptane which is an original constituent of the crude benzole, is more stubborn, and defronting seems to be only a modest help.

(E) Future Developments Towards Higher Quality and Higher Yields

Although, with present British and Continental specifications for pure benzene, a traditional distillation as described previously seems quite economical on yields and working costs, further improvement, both in quality giving of pure benzene of 99·9 per cent purity and 99 per cent yield, could well be effected by either solvent extraction (both Udex and Lurgi have effective plants) or by the "freeze" technique of pure benzene developed by Newton Chambers.

Azeotropic distillation could also be economically operated with the low concentration of non-aromatics, and has been successfully used, although it is felt that solvent extraction or "freeze" techniques would be more economical in both lower working costs and higher yields.

Thus, any further developments in specification requiring crystallizing-points of the order of 5·45°C upwards can easily be met at even higher yields of pure benzene, but, with little present demand for such a specification at the right price, these developments do not at present justify the additional capital expenditure and additional working costs.

(F) Stripping of Hydrorefining Plant Residue

The residue from the hydrorefining plant, about 4 per cent, v/v, of the crude benzole, has an approximate composition as follows :—

	Per cent.
Boiling up to 155°C	50
Boiling between 155° and 190°C	25
True residue	25

This has to be stripped, recovering approximately 75 per cent of material to 190°C as recycle to the crude benzole, and a residue being about 50 per cent naphthalene.

(5) THE USE OF COKE OVEN GAS AND ITS VARIATION OVER THE PLANT

(A) Discussion

By utilizing the hydrogen content of coke oven gas, the problem of the expense of hydrogen is small. The problem is, therefore, one of utilizing a crude gas for a chemical usage, and the designing of a robust plant capable of long on-stream times.

Coke oven gas supplies both the hydrogen and the heat to the process. Although it is desirable to have as high a hydrogen content as possible, in practice it is taken as made at the coke ovens. A normal coke oven gas has a hydrogen content of 55 to 62 per cent, v/v, so that, to maintain a partial pressure of 18 atm at the catalyst, this must necessitate not only a constant bleed-off of gas, but also running the plant at a higher pressure than when using pure hydrogen. This pressure is normally in the range of 40 to 50 atm.

It is also necessary to ensure that the oxygen content of coke oven gas is low, and that trace impurities are eliminated. For this reason, the gas at Port Talbot is pre-treated between the second and third stages of a four-stage fresh-gas compressor.

Coke oven gas is also used as the heating medium, as the low bromine numbers of crude coke oven benzoles do not supply sufficient heat of reaction for the heat needs of the process. The additional heat is supplied by the catalytic burning of the tail gas by controlled air flow, and this accounts for the nitrogen in the final gas.

(B) Analysis of Fresh and Final Gas With Intermediate Tail Gas

Over the guarantee run, which is typical of normal running, the following were the compositions of fresh and final gas.

Fresh gas	= coke oven gas as received.
Tail gas	= expansion gas let down 80 lb/in² gauge
Blow-out gas	= release from hydrogen sulphide blow-out tower at 70 lb/in² gauge.
Final gas	= total outlet from the plant.

The final gas equals the tail gas after the catalytic burning, plus the blow-out gas (which does not go through the catalytic burner).

On a throughput of 627 gal/h of crude benzole, the observed spot gas analyses were as shown in Table 15.

Table 15.

	Fresh Gas	Final Gas	Tail Gas
1	2	3	4
Gas Quantity (s. ft³/h) ..	20,723	14,505	12,300
Composition (v/v) :—			
CO_2	1·8	2·0	0·6
C_nH_m	1·0	0·4	0·4
O_2	0·0	0·0	0·0
CO	6·0	0·8	1·4
H_2	60·0	29·7	46·8
CH_4	27·3	51·2	50·7
N_2	3·9	15·9	0·1
H_2S (gr/100 ft³)	260	1,360	Not taken
B.t.u./s. ft³ (net)	442	565	590
(gross) ..	501	612	665
Specific gravity	0·340	0·507	Not taken
Added air recorded at 2,650 ft³/h.			

As these are plant metered figures and analyses on spot samples only, it would be unwise to draw arithmetical conclusions from them. If the figures are taken at their face value, there are some obvious discrepancies of quantitative balance due to samples being taken at different times. However, they do give some good indication of the change of composition of gas over the process, and certain observations can be usefully made.

(i) The increase in hydrogen sulphide content of the gas is great and is, in fact, a function of the sulphur in the crude benzole, which all appears as hydrogen sulphide after hydrogenation.

(ii) The quantity of final gas is less than the quantity of fresh gas. In this case, it is 70·4 per cent, v/v. As 15·9 per cent, v/v, of the final gas is nitrogen introduced as air in the catalytic burning of tail gas, the true tail gas without the nitrogen diluent is 57·8 per cent, v/v, of the fresh gas.

(iii) It is the hydrogen content of the coke oven gas which is effectively used both for process hydrogen and heating. The latter is not only expected from a typical reaction of this kind, but is confirmed by the difference in hydrogen content of tail gas (before catalytic burning) and hydrogen content of final gas. It is estimated later that two-thirds of the hydrogen in the fresh coke oven gas is used.

(iv) Although there is a decrease in quantity of tail gas compared to fresh gas, there is an increase in its calorific value per s.ft³, despite the dilution due to the nitrogen content of air used in the catalytic burning. This is because there is a decrease in quantity of low-calorific value hydrogen, and an increase in quantity of methane due to the products of hydrogenation.

(v) On the figures quoted, the thermal usage per hour as gross calorific value is :

	Therms/h
Fresh gas 20,723 s.ft³ 501 Btu/ft³ ..	103·82
Final gas 14,505 s.ft³ 612 Btu/ft³ ..	88·77
Overall usage	15·05
On 627 gal/h	2,400 Btu/gal.

This figure, or slightly less, is being regularly returned on monthly plant running performance.

(6) HYDROGEN USAGE ACTUAL AND THEORETICAL

(A) Actual Usage

By using the figures obtained in the method shown previously, the hydrogen, methane and carbon monoxide variation per hour are as in Table 16.

TABLE 16.

	Fresh Gas		Final Gas		Difference
	20,723 s.ft³/h		14,505 s.ft³/h		6,218 s.ft³/h
	Per cent (v/v)	s.ft³	Per cent (v/v)	s.ft³	s.ft³
Hydrogen ..	60·0	12,434	29·7	4,308	−8,126
Methane ..	27·3	5,657	51·2	7,427	+1,770
Carbon monoxide	6·0	1,243	0·8	116	−1,127

Hydrogen in tail gas as percentage of hydrogen in fresh gas, 34·6 per cent, v/v.
Methane in tail gas as percentage of methane in fresh gas, 131·3 per cent, v/v
Carbon monoxide in tail gas as percentage of carbon monoxide in fresh gas, 9·3 per cent, v/v.

From the figures, on their face value, about two-thirds of the hydrogen in the fresh gas is used, but there is a gain equivalent to one-third of the methane in the fresh gas from the reaction products of hydrogenation, in particular, it is suspected, the hydrogenation of carbon monoxide to methane. However, these figures should be taken with the reservation of accuracy that they are calculated from plant meter figures and spot gas analyses.

(B) Theoretical Usage of Hydrogen Based on Feed Composition

In the guarantee run, to which the gas analysis refers, the composition of the crude benzole feed was as follows :—

Specific gravity	0·887
Total sulphur, w/v (per cent)	0·44
as thiophen, w/v (per cent)	0·23
as CS_2, w/v (per cent) ..	0·21
Bromine No. (g/100 ml) ..	13·4
Pyridine Bases	Not observed ; estimated to be very low.
Naphthalene	Not observed ; estimated to be about 3·2 per cent.
Naphthalene hydrogenated, estimated on tetralin produced, 1·4 per cent	

Using the formula previously established, the hydrogen requirements per hour on throughput of 627 gal of crude benzole are as shown in Table 17.

TABLE 17.

	Per 1,000 gal (s.ft³ of Hydrogen)	Actual Content of Crude Benzene	Hydrogen Usage (s.ft³/h on 627 gal/h Crude)
1	2	3	4
0·1 per cent, w/v, sulphur as CS_2	225	0·21 per cent, w/v	296
0·1 per cent, w/v, sulphur as thiophen	450	0·23 per cent, w/v	649
1 g Bromine/100 ml ..	225	13·4	1,890
0·1 per cent, w/v, naphthalene to tetralin	56	1·4 per cent, w/v	492
			3,327

On a throughput per hour of 627 gal, the crude benzole would require 3,327 s.ft³/h of hydrogen or 5·3 s.ft³/gal of crude.

(C) Hydrogen Used for Heating

In the catalytic heater, assuming that only hydrogen combines with oxygen and is converted to water in the catalytic burning, 2,650 s.ft³/h of air equals 530 s.ft³/h of oxygen, and 2,120 s.ft³/h of nitrogen.

Secondly, to verify this the nitrogen content of the final gas is 15·9 per cent, v/v, of 14,505 s.ft³/h, which is equal to 2,306 s.ft³/h of nitrogen.

This is within the range of anticipated verification we can expect. Therefore, using the first figure, 530 s.ft³ of oxygen combines with 1,060 s.ft³ of hydrogen in the catalytic burning to provide heat. This is an assumption.

(D) Carbon Monoxide Conversion to Methane, Using Hydrogen

Assuming that this is hydrogenated to methane and water, 3 ft³ of hydrogen being required for each 1 ft³ of carbon monoxide, 1,127 ft³/h or carbon monoxide would combine with 3,381 ft³ of hydrogen.

(E) Summary of Hydrogen Usage

	s.ft³/h.	Per cent.
Apparent hydrogen usage.. ..	8,126	
For hydrogenation of crude benzole	3,327	40·9
For carbon monoxide conversion	3,381	41·7
For catalytic burning	1,060	13·0
Unaccounted-for hydrogen ..	358	4·4
	8,126	100·0

In drawing this conclusion of a hydrogen balance, it is based on certain assumptions that are not necessarily correct, and based on plant meter figures and spot analyses of gas that have limitations as to accuracy. A different set of assumptions on gas reactions could lead to a differeng conclusion, although it does seem that the assumptions made fit the figures recorded.

(F) Carbon Dioxide

The figures recorded make little sense as there seems to be a large decrease of carbon dioxide over the process. However, on checking regularly recorded figures of fresh and tail gas analyses, it would seem that the carbon dioxide content of the fresh gas is concentrated in the tail gas ; thus, on this analysis, one would normally expect the carbon monoxide in the tail gas to be about 2·5 per cent, not 0·8 per cent recorded. As the tail gas is in contact with the water effluent before it is withdrawn at a pressure of 150 lb/in² gauge and about 50°C, one would expect some carbon dioxide to be withdrawn here, but as the quantity of water effluent is small it would not exceed 300 to 400 ft³/h.

There is, therefore, enough evidence to support the theory that there is no appreciable make or loss of carbon dioxide over the process, which in turn would support the theory that carbon monoxide is converted to methane.

(7) ECONOMICS OF HYDROREFINING COKE OVEN BENZOLE WITH COKE OVEN GAS

(A) Throughput and Capital Cost

The larger the unit of throughput, the more economic does hydrorefining become, compared to acid washing, both in working costs and comparative capital cost. From throughputs of 10 mill. gal/a upwards, hydrorefining becomes an obvious choice, and for large centralized units as are the practice in the United States of America and Europe the appeal of hydrorefining is very strong.

However, in the United Kingdom our commercial pattern is of many small units usually below 10 mill. gal/a, and it is on one of these small units that the economics are worked out, namely, 4·8 mill. gal/a of crude benzole. The capital cost of the hydrorefining section only is approximately £260,000. The total cost for a whole works built on virgin land, including hydrorefining, distillation, one month's storage, laboratory, administration block, all pipework and site preparation would be about £500,000.

(B) Cost of Running 16,000 gal/d Plant

Firstly, let us look at the economics of running the hydrorefining section only, the refined product requiring further distillation, and the crude benzole being to the following characteristics :—

	Per cent
Benzole Producers Evaluation water white to 160°C	88·2
Creosote residue	7·9
Loss allowance	3·9
Total sulphur (w/v)	0.44
as carbon disulphide (w/v)	0·21
as thiophen (w/v)	0·23

TABLE 18.

Throughput : 4·8 mill. gal/a.

Item	Unit/gal of Feed	Cost of Unit	Cost/gal of Crude (Pence)	£/a
1	2	3	4	5
Labour	1 man/shift	£3,600/year	0·180	3,600
Steam	1·4 lb./gal	100*d.*/1,000 lb	0·140	2,800
Coke oven gas	2,400 Btu/g.	8*d.*/therm	0·192	3,840
Purification Final Gas	—	Say £2,000	0·100	2,000
Electricity	0·4 kW/g.	1·1*d.* per kW	0·440	8,800
Cooling water	15 g./g.	4*d.* per 1,000 g.	0·060	1,200
Condensate	10 g./100 g.	2/6*d.* per 1,000 g.	0·003	60
Catalyst replacement (25 per cent/a)			0·100	2,000
Maintenance : say £10,000			0·500	10,000
			1·715	34,300
Overheads :				
Depreciation (10 per cent of capital cost)			1·300	26,000
Works overheads (viz., supervision, insurance, rent rates, laboratory) : say			0·500	10,000
		Overhead Costs	1·800	36,000
		Total Works Costs	3·515	£70,300

Note:—No charge for interest on capital.

This, compared to acid washing costs, not including distillation, on the same basis, comes out at :

	Pence/gal.
Prime costs	1·5
Overheads including depreciation	0·7
Total working costs	2·2

(C) Comparative Revenue and Costs for Hydrorefining and Acid Washing a Similar Crude Benzole

To compare acid washing with hydrorefining costs, we have to compare yields as well as costs, based on throughput of 4·8 mill. gal/a, which is shown in Table 19.

TABLE 19

Revenue/100 gal :—

	Hydrorefining		Acid Washing	
	Per cent	Pence/100 gal	Per cent	Pence/100 gal
1	2	3	4	5
Forerunnings @ 1/0*d.* per g.	Nil	Nil	1·5	18
Pure benzole, toluene, xylene, @ 43*d.*/g.	85·9	3,694	81·7	3,513
Motor benzole @ 33*d.* less 1*d.* low gravity penalty 32*d.*	4·0	128	Nil	Nil
Solvents above xylole 30*d.*	2·6	78	2·6	78
Creosote residue 8¼*d.*	3·2	35	8·2	68
Physical loss	3·3	Nil	6·0	Nil
	100·0	3,935	100·0	3,677
Average realisation per gal of crude benzole		39·35*d.*		36·77*d.*
Prime Costs per Gallon :		*Pence*		*Pence*
Hydrorefining		1·72		—
Acid washing		—		1·50
Distillation related per gal of crude benzole		2·00		1·90
Storage, handling, yard, *etc.*		1·00		1·00
Overheads :				
Administration charges : say		1·50		1·50
Plant depreciation at 10 per cent :				
Capital cost, hydrorefining and distillation, £350,000		1·75		
Capital cost, acid washing and distillation, £200,000				1·00
Other Depreciation at 5 per cent:				
Capital cost storage, services, piping, civil engineering, £200,000		0·50		0·50
Total costs per gal of crude		8·47		7·40
Net revenue per gal of crude		30·88		29·37

Advantage for hydrorefining against acid washing : say 1·50*d.*/gal.

Note:—No inclusion for interest on capital.

(D) Conclusion

It is therefore apparent on this smaller-scale operation that, although the net revenue is greater by hydrorefining by about £30,000/a, the return on capital using a 10-year depreciation is slightly lower. However, the additional expenditure of £150,000 on a hydrorefining plant gives £30,000/a additional revenue. However, this is comparing a yield and cost to No. 2 pure benzole specification, and as this specification becomes tighter the yield on acid washing would drop slightly and the costs increase.

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PRESENTATION, DISCUSSION AND REPLY AND VOTE OF THANKS

Presentation

Mr. J. B. Lane presented his paper and summarized its contents.

Discussion

Mr. F. K. Illingworth (Clitheroe) :—When the scheme of hydrorefining was put forward at Cadishead, we in Lancashire had to give much thought to it, and I think it was fairly unanimous that the right decision was being made, certainly with the naphthas that had to be processed. One striking feature in the paper is the Author's reference to the fact that products are being made ahead of specification. So often one is chasing specifications and making products in arrear. With this plant, Mr. Lane is making products ahead of specification.

I have inspected the Cadishead plant on a number of occasions, and only a few days ago, when I saw the mess underneath our old benzole washer, I said : " There is only one man correct around here, and he is Joe Lane, who has no acid tar to deal with "!

The Author refers to a catalyst that might conceivably convert paraffinic materials into aromatic bodies as desired. Is it possible for him to consider transferring the aromatics from one group to another ; in other words, could this be adapted to converting toluene to benzene, or *vice versa* to meet market demands, because my experience of benzole refining is that one cannot sell all products evenly?

With regard to the flow diagram, I am intrigued by the primary cooling of the product by the direct injection of water. Is there any reason for this? Undoubtedly, it is a most efficient way of cooling. Would the Author comment on why this method was chosen by the designers?

Having started off, at Cadishead, with hydrorefining, the Author went to acid washing at Scunthorpe, and, in spite of his alteration in the dates in the paper, what intrigues me is the fact that, when he came to the third decision, he made it before either of the other plants was in operation. Port Talbot was conceived in March, 1959 ; Cadishead was started in June of that year, and Scunthorpe in about August. The fact that hydrorefining was chosen a second time before any plant was in operation is a measure of Mr. Lane's confidence.

It has to be admitted that Mr. Lane is one of the few people who are in a real position to compare costs ; therefore, his Table 19 must command very great respect indeed, and also his indication that the advantage of hydrorefining as opposed to acid washing is 1½*d.*/gal. In my opinion, this statement alone indicates the death knell of acid washing so far as crude benzoles are concerned.

Mr. W. J. Pater (British Coke Research Association) :—I should like to congratulate the Author for a well set-out paper giving many valuable details. I am also impressed by the way the arguments for and against the selection of hydrorefining or acid washing are presented in the paper. Dealing with some of the points arising from the paper, I should like to ask : Why was hydrorefining adopted at Port Talbot although the throughput of benzole was less than 6 mill. gal/a? Was the difficulty of effluent disposal mentioned by the Author the only reason?

In the discussion of the paper by Bradley and Middleton, as well as in the Author's presentation, a suggestion was made that in the near future the specification for synthesis-grade benzole would be more stringent. Can this specification be met after hydrorefining by fractional distillation of crude benzole of low paraffinic content? Under this condition, is hydrorefining still the preferable method, or would the paraffins formed during the treatment cause difficulties in the final fractionation?

It would seem that the hydrorefining process is the method of the future because, according to the Author, amongst other merits, it affords in general the more economical way of producing benzole of high yield and purity.

I have two further queries : one of those mentioned already by Mr. Illingworth, refers to the method of cooling the gas. I notice that, after the final heat exchanger, water is added directly to the gases, followed by an indirect cooler. What advantage is there in the use of both direct and indirect cooling, and could the latter be used, as it avoids the production of effluents?

The second question is this : In view of the experience the Author had with both pure hydrogen and coke oven gas, at what ratio of prices between pure hydrogen and coke oven hydrogen would he be prepared to recommend coke oven gas? Further, what, in his opinion, would be the minimum acceptable percentage of hydrogen in coke oven gas?

Dr. A. E. Haffner (Southern Gas Board ; Senior Vice-President, The Institution of Gas Engineers) :—Most of us in the gas industry do not appear to be directly concerned with the problems that have been exercising Mr. Lane's mind. We have, therefore, been watching from the edge of the field the game he has been playing and have seen some of its finer points.

The common fields of interest between the coke oven industry and the gas industry certainly include that of obtaining higher financial yields from benzole. The changing nature of the gas industry brings up the subject of hydrorefining in a number of contexts. Some of us have come across it because the refinery gas that we may buy can have been derived from several processes on a refinery, including that of the hydrorefiner as well as the catcracker. It might also be valuable for the gas industry to hydrorefine some of its raw materials, liquid and gaseous, to enable them to be reformed subsequently in continuous catalytic processes by using some of the final hydrogen-rich gas. Many more of us are interested in the possibility of up-grading benzole to ensure an adequate return from the already installed benzole plants, which are still virtually necessary from the point of view of associated organic sulphur removal from the gas, but it is precariously placed from the point of view of preference and when its costing in the gas works is based on accountants' views about replacement therms.

The work described by the Author is, therefore, important because it provides valuable information on which to base decisions as to whether benzole hydrorefined in suitably large units based on modern chemical engineering can command a market price in the form of a monetary or physical premium such as will make it attractive to continue to extract benzole in our changing industry.

Mr. H. J. Victory (Port Talbot) :—I should first like to congratulate the Author on this paper, which I think is historical as representing perhaps not a break-through, but a very big break into the field of hydrogenation products. It is also historical because one has never previously heard Mr. Lane confess to nervousness. I think there is no doubt that the paper will be sought for some years to come as a valuable work of reference.

The number of data in the paper are so great that I have had difficulty in digesting them, let alone commenting on the paper in the time available, but there are a few minor points I should like to raise.

In Table 7, I am puzzled by " Condensate 5 gal/100 gal feed : 2*s.* 6*d.*/1,000 gal." I am not sure whether this refers to effluent or to something else : perhaps the Author would explain what it is.

The other point that interests me is the method of pressure let-down of the raffinate and release of expansion gas. It is described at Cadishead as being let down initially to 5 atm, and at Port Talbot at 150 lb for about 10 atm in two stages. I suppose this pressure is calculated to prevent evaporation and loss of desired constituents of the liquid raffinate. Could the Author explain why these particular pressures were chosen, and why there is a two-stage let-down at Port Talbot?

At the Port Talbot plant, a recycled gas, after being mixed with vaporized feedstock, is passed through the " pre-catalyst ". I wonder what the pre-catalyst is, and what it does. The Author describes coke oven gas as passing through the scrubber, and I wonder what the gas is scrubbed with and for.

The Author then mentions that fresh coke oven gas passes through pretreatment to remove oxygen and other undesirable constituents. Would he say what this pretreatment is, and what is removed besides oxygen? I am rather sensitive about the quality of my coke oven gas.

The benzole feedstock is quoted at 0·21 per cent of carbon disulphide, and according to Table 17 this accounts for between 8

and 9 per cent of the hydrogen demand of the process. I wonder whether there is any case for incorporating any form of defronting to take out this carbon disulphide. It seems from the diagram that there might be enough heat left in the reactor gas, after passing the evaporator, actually to do this.

Mr. H. C. Wilkinson (British Coke Research Association) :—During the last decade, there has been an outstanding growth in the amount of pure benzole required by the chemical industry. It is stated that it has increased from 10 mill. gal in 1950 to 40 mill. gal in 1961, and will increase to 70 mill. gal in the mid-60s. This growth has occurred because of the large increase in the manufacture of chemical products such as nylon, polystyrene, and analine which can readily be produced from high quality aromatic hydrocarbons.

The coke industry has long been the source of these valuable materials. Great efforts have been made to improve the crude benzole, principally by acid washing techniques, but now the alternative process of hydrorefining offers an attractive way to improve quality. It would appear that, if the present trend continues, it is unlikely that much more crude benzole can be expected from the coke oven industry ; therefore, it is highly likely that the gap between production and demand will be filled by refining products from the petroleum industry, using such techniques as catalytic reforming. It is fair to say that the benzole produced from the coke oven gas is an attractive compound and has advantages over a similar product produced from crude petroleum, because crude benzole produced by coal carbonization contains a much larger quantity of aromatics than crude petroleum ; even after processing by the techniques mentioned, refined petroleum rarely contains more than 60 per cent of aromatic material. To produce a final product to the desired specification involves the use of more plant and greater expense.

The Author has mentioned in Table 19 that hydrorefining yields approx 85 per cent of material boiling below 160°C, and I should be interested to learn whether he has any figures available for the percentage of benzene, toluene and xylene in this product.

Hydrorefining has other attractions : of automation, of economy in use, of reagents, and virtually no effluent problem. Therefore, it appears that crude coke oven benzole refined by the hydrorefining technique offers a unique source of valuable materials likely to be required in ever increasing quantities.

The Author has mentioned that the capital costs of hydrorefining plant and its associated plant are considerable, but this should not be allowed to deter one from considering the construction of a number of plants. Coke manufacture and coke ovens are broadly confined to three main areas—Durham, South Yorkshire and the North Midlands and South Wales—and it might be suggested that central refineries could be constructed to refine the by-products from the coke oven plants in those districts.

Mr. C. Ryder (Central Laboratories, South Eastern Gas Board) :—Mr. Lane's paper is very timely, coming, as it does, when the whole future of benzole extraction and purification is being reconsidered because of the uncertainty in demand and future price realization.

As the Author explains, the specification for No. 2 pure benzene is not a very rigid one, and entry into the European Common Market or the European Free Trade Area might force a much more stringent specification to be applied. It would be very helpful if this future demand could be much more clearly defined both in terms of output and of purity to enable forward planning of new projects to be carried out with confidence.

A joint committee of the North Thames and South Eastern Gas Boards has been looking at alternative methods of upgrading crude benzole ; as with Lancashire Tar Distillers, Limited, there exists the problem of what to do with the low-gravity vertical retort naphthas. Mr. Lane's paper has certainly helped to clarify the first stage, that is, removing sulphur compounds and stabilizing the naphtha, but the problem still exists as to how to separate the aromatics from the paraffins.

It should be remembered that, even if separation proves to be simple, the value of the paraffins will be very low, being similar to that of light distillate feedstocks used for gas making ; hence, unless a very high price can be obtained for the aromatic hydrocarbons, the cost of the plant and its operation must be very cheap.

In various places in the paper, there is reference to the sulphur content of the raffinate being below 1 p.p.m. I should like further clarification of this point. Do the sulphur contents in this paper refer to the refined benzene fraction alone, or to the product as it leaves the hydrorefining stage? If the latter, are there times when the purity exceeds this figure, and, if so, by how much? It might perhaps have been better if I had prefaced these remarks on sulphur content by asking whether the present method of estimating sulphur is very accurate, particularly in view of the recent Benzole Producers' research paper on the Rhone-Poulenc method.

The sulphur content of the hydrorefined naphtha is of importance because subsequent treatment of the naphtha may involve dealkylation in one way or another, and the methyl thiophens, *etc.*, may appear as thiophens in the final product benzene. Assuming one had a hydrorefined benzene containing a few parts per million of sulphur, what additional steps could be taken to reduce the sulphur content below 1 p.p.m.?

It is noted that a cracked petroleum feed led to trouble in the hydrorefining plant. Has the Author been able to elucidate what caused the trouble ; olefines, perhaps? This point is of importance where light distillate is used on water gas plants and is not fully cracked, because if benzole extraction is practised on the mixed coal gas/water gas streams, a potential source of trouble could lie in the processing of the benzole.

Has the Author any knowledge of the relative activities of the catalyst to thiophen and its homologues, and can the activity be varied by altering the composition of the catalyst?

Is the Author aware of any other disadvantages due to the presence of carbon monoxide in the process gas, other than the hydrogen-consuming methane synthesis reaction?

With reference to plant operation, how is polymerization in the evaporator controlled at Cadishead? The paper notes that cleaning at six-monthly intervals is necessary at Port Talbot. In both the diagrams of Cadishead and of Port Talbot, hydrogen or coke oven gas is fed direct to the reactor. Is there any advantage to be gained by mixing all the hydrogen with the feed before the evaporation? Lower partial vapour pressures might lead to lower evaporator temperatures and less polymerization. Of interest is the six-stage evaporation of the feed. Why is this necessary?

A final word regarding size of unit : Why does the Author state that 10 mill. gal/a is an economical size. Is it because of reduced capital cost relative to scale, which is said to come with increasing size? ; or is it lower operating costs and overheads?

Mr. D. Bradley (Newton Chambers and Company, Limited) :—The wide variety of aspects dealt with in the paper is quite extraordinary, and I think that the Author has contributed to the literature on hydrorefining by putting together a collection of information such as has not hitherto been available. It is indeed refreshing to read a paper on benzole refining at a time when preference appears to be the main concern of everybody in the benzole industry.

My feelings on the more general comments made in the paper are naturally in sympathy with those of the Author, but I differ in some of the more detailed points. The experience of Newton Chambers in operating what might better be described as a small industrial unit would substantiate the Author's comment regarding the ease of operation of the hydrorefining plant. An operator who came to the company at the age of 60, having had no plant experience whatever, was capable of operating in the capacity of shift attendant, such was the simplicity of operation of the unit.

The company would corroborate the Author's statement regarding the increased yield to be obtained when operating a hydrorefining plant as opposed to an acid washing plant, and its figures indicate a 20 per cent increase in recovered toluene and 100 per cent increase in recovered xylene. The benzene yield is not considerably increased.

Unfortunately ,the figures I have quoted are of a comparison made between hydrorefining and batch acid washing, which is significant because, if the figures were available for continuous acid washing, the difference would not be quite so much in favour of hydrorefining. It has been pointed out that it is possible by distillation to recover high yields of high crystallizing-point benzene from a crude coke oven benzole. In point of fact, in the company's experience this is true, and it feels it should be pointed out that only one distillation stage is involved when distilling hydrorefined benzole, whereas in the more normal practice two stages are involved when distilling acid-washed benzole. I think that the

struggle over the past five or 10 years for hydrorefining or acid washing to assert themselves is still raging. It appeared that hydrorefining was well ahead on points at one time, but recent successes of continuous acid washing appear to have evened up the contest. As the Author puts it, it is up to the individual operator to make his own assessment. Each case must be treated on its own merits.

Mention has been made of the C.T.R.A. process for naphtha refining. Newton Chambers has processed vertical retort naphtha similar to that at Cadishead. Hydrorefining was carried out at a pressure of 300 lb/in² with hydrogen content in the equilibrium gas over the reactor of 35 per cent. Acid washing with 65 per cent acid was carried out, and the products were water white colour stable naphthas. The degree of hydrogenation was not so great when using pure hydrogen at 600 lb/in². The advantages with the C.T.R.A. method of operation are that the high yield claimed for any hydrorefining process is still obtainable, yet there is a lower hydrogen consumption. Also, the residual compounds are probably better solvents than the corresponding paraffins, and the naphthas contain aromatics and retain their flatter smell.

I do not wish to indulge in a discussion on the commercial cost assessment in the paper. There is, however, one point that I should like clarified in the cost assessment, that is, the reason for the omission of the charge for the interest on capital. I should also like to ask whether the Author has any information on the possible increase of normal heptane during hydrorefining. In Table 10, the normal heptanes are together with other hydrocarbons.

The Author has mentioned the possibility of benzene refining by the crystallization process. This process was developed by Benzole Producers, Limited; C.T.R.A., and Newton Chambers, and Newton Chamber has carried out work on upgrading paraffinic feedstocks. This work has now been concluded, and commercial plant can be offered. The removal of impurities other than thiophen can be carried out with this process, giving high yields of benzene at low operating cost. At the instigation of the Author, the company has recently been investigating the application of the crystallization technique to reduce the thiophen content of benzene, in which the thiophen content has already been reduced to a very low level. The results to date have been encouraging from a scientific, although as yet not necessarily from a commercial, point of view, and an approximate idea of the degree of upgrading may be gathered from the fact that in the single-stage crystallization process the thiophen content of the product is 50 per cent of that in the feed with a 50 per cent yield obtained. Further development work is progressing along these lines. In a different context, it has been demonstrated that thiophen can be removed from crude benzene fractions, and where these occur when the sulphur content is of the order of 0·28 per cent or below, then high yields of No. 1 pure benzole may be produced without acid washing.

Mr. L. G. Townsend (Liverpool) :—This paper is a very welcome and useful addition to the knowledge of this process, which has been slowly accumulating over the last four of five years. In many directions, the results go much further than previous papers in that detailed costs of actual operation in this country are given for the first time, not only when treating coke oven benzole, but also when dealing with the more paraffinic tar naphtha. The conclusion in the early days of hydrorefining, that such plants must be of relatively large size to become economic, has been confirmed, but this is an advantage rather than otherwise at the present time. The number of hydrorefining plants is steadily increasing from the mere three that were at work as recently as 1956. The details given in the introduction to the paper regarding the data on which the decision to install the Lurgi plants was based are of intense interest, and it is quite evident that considerable courage was needed to come to the decision to install a hydrorefining plant at Cadishead.

The gas industry's interest in benzole is somewhat cursory at the present time, and it seems unlikely that there will be a revival of interest in the immediate future. It is possible, however, that the renewed interest in smokeless fuel might eventually lead to some renewed interest in benzole recovery, although the impact of the Common Market would probably be decisive in the issue. The manufacture of reactive smokeless fuels, particularly of the premium type, does, however, lead to the production of tars of a progressively paraffinic type, and this could well react unfavourably on the hydrorefining process. It seems to me that the real advantages of hydrorefining tend to be lost with feedstocks of mixed hydrocarbon types. It would also seem that the optimum situation for a hydrorefining unit is in close proximity to large coke oven works recovering benzole and furnishing a supply of neat coke oven gas. It is unnecessary to say anything further regarding the manifest advantages of hydrogen treatment as opposed to acid washing, but there is no doubt that one of the major advantages lies in the preparation from aromatic feedstocks of products of extreme purity.

There are a number of questions that occur to one when studying the paper, and I should like to ask the Author a few of what seem to me to be the more important ones. When one compares the cost of the process at Cadishead and Port Talbot, one is immediately struck by the advantage enjoyed at the steelworks, which amounts to nearly 2¼*d.*/gal of feedstock. This is due in part to the differences in the feedstock leading to higher hydrogen consumption, but there would appear to be at least 2*d.* additional cost due to using pure hydrogen as against coke oven gas, even allowing for compression charges. It would be interesting to know whether the production of hydrogen in an independent plant would not be justifiable in such a case?

There are one or two points that might possibly have been of interest during the annual overhaul of the plant. How often is catalyst regeneration necessary, and what are the main factors influencing this? During the last few years, one has heard a great deal about hydrogen sulphide corrosion of steel where catalytic reforming is carried out at pressures above 30 atm and temperatures over 300°C. This can result in heavy scaling and sometimes in inter-granular cracking, and alloy steels are sometimes essential. Are any difficulties of this kind apparent in the plants described? It is evident, when dealing with benzole as the feedstock, that some kind of pretreatment is necessary to avoid the formation of resinous materials from the diolefines. Would it be profitable to recover these diolefines, which are in steady demand, by means of a preliminary defronting of the crude benzole? Presumably, no difficulties of this kind arise when coal tar naphthas are the feedstock?

It is interesting to note that when using coke oven gas there would appear to be a loss of carbon dioxide in this process, and one can only assume that the water effluent is responsible for the removal. I recall that the conversion of carbon monoxide to methane has been established as a common mechanism in the hydrorefining process. Indeed, it has been suggested that the preliminary removal, or conversion, of carbon monoxide is worth while in that it reduces hydrogen consumption. It is somewhat ironical that the very reaction that was studied in detail in the early days of the former Gas Research Board with a view to synthesizing methane should be an embarrassment in the present process. Whilst on the subject of hydrogen consumption : it has always seemed to me that it might be possible to recover more naphthalene from the tar oils by refrigeration, and this would be of some assistance in reducing unprofitable hydrogen usage. I should be glad to know what the Author thinks about this aspect.

I am interested to see that the complete removal of hydrogen sulphide from the hydogenated products is achieved by pressure let-down. I should have expected a small alkali wash would have been necessary, since hydrogen sulphide seems to have a small solubility in almost all organic solvents. I should be interested also to know the method used for removing the oxygen from the coke oven gas, since, with a crude gas containing hydrogen sulphide, catalytic removal might be possible using iron oxide.

In conclusion, I must congratulate the Author in preparing a paper that is full of valuable information on a most interesting process.

Mr. Thornton Scott (Sheffield) :—With regard to basic loss during acid refining, Mr. Bradley has said there is no up-to-date acid washing plant in this country. The major loss in the refining is due to the filling of vessels. There are five fillings involved in the operation, and I suggest that, if crude benzole were put through the operation without any acid washing whatever, losses between 2 and 4 per cent would be realized at the finish.

I should like to congratulate the Author on the factual representation of his comparisons. In a paper that I have recently read in The Coke Oven Managers' Association's Yearbook, a yield was accepted of 97½ per cent for hydrorefined products, and a yield of 78 per cent was assumed in the case of acid washing ; so the figures appeared more favourable in that paper.

The unsaturated heavy solvent naphthas in Table 19 ought, I think, to carry a slightly higher price.

Mr. G. Claxton (Benzole Producers, Limited) :—I, also, should like to congratulate the Author on giving a comprehensive account

of the hydrorefining plant at Port Talbot. Let me say straight away that hydrorefining is not new, but for many reasons failed to make headway in Great Britain until the Lane brothers went ahead with the two plants described, and with which they are obviously not disappointed. I think I am right in saying that I was one of the early persons in this country to see hydrorefining in 1939 : it was in Berlin. Although the plant had many shortcomings, it was a beginning but my favourable report proved to be a cry in the wilderness. I saw also the first Lurgi plant, and I think that the report produced had something to do with persuading the Lancashire Tar Distillers Company to install the first plant in Great Britain.

Whilst I can see many advantages in hydrorefining, I think my views have changed a little since 1955, due, particularly, to improved techniques that have made acid refining more competitive. Indeed, I am sure that the Author has painted an unduly pessimistic picture of acid refining. If we look at Table 13 in conjunction with Table 10, we can see that the crude benzole contains hydrocarbons up to the xylenes, 91·8 per cent of aromatics, 4·3 per cent of paraffins below benzene and olefines, and 3·9 per cent of material above xylene. By hydrorefining, about 98 per cent of the aromatics have been recovered, whereas by acid washing only 89 per cent are accounted for. Somewhere, 8 per cent of the aromatics have vanished. In the original crude, residues amount to 8 per cent, whereas after acid refining these constituents add up to 15·7 per cent. In other words, the missing 8 per cent has turned up either as polymer or loss. Whilst I agree that during acid washing some of the aromatics do vanish, I find it difficult to believe that the loss is as great as that suggested.

In Table 19, the cost given for acid refining plant seems high compared with the hydrorefining plant, which includes a highly efficient fractionating plant. I am sure that by using continuous acid refining plant and by adjustments in cost an equally attractive picture can be made for acid refining. I do not say this to denigrate hydrorefining, which has many advantages, one of which is the more ready disposal of effluent. I do say, however, that, given sufficiently good crude benzole, free from aliphatics and low in thiophen sulphur, it is probably easier and cheaper to prepare pure products by acid refining. For crudes containing aliphatics or high sulphur, however, the advantages of hydrorefining increase. In these circumstances, I should not rely on super-fraction for separating pures, but would install freeze refining.

In the future, there is likely to be a good market for high quality benzene, but with poorer prices for toluene and xylene. Given large enough plant, hydrorefining followed by freeze refining holds its own, but for smaller units acid refining will probably still win.

A question was asked earlier, whether it would be possible to demethylate the toluene and xylene. Plant is being offered in the United States that, by the use of catalysts, is said to do exactly this by recycling toluene and xylene. The methane formed can be reformed into hydrogen and supply the hydrogen needs of the plant, which is of some value to a tar distiller who is not near to a coke oven. It is rather significant that, in the United States, the two large hydrorefining plants I have seen both produce hydrogen, although they have coke oven gas available. One is producing hydrogen from natural gas, and another from ammonia cracker.

Mr. J. F. Ritson (Simon-Carves, Limited) :—I was interested in the problem of the separation of paraffins formed during the hydrorefining stage. Quoting from an earlier paper, by Bradley and Middleton, four important compounds are : *cyclo*hexane, b.p. 80·9°C ; 2 : 4 dimethyl pentane, b.p. 80·7°C ; *n*-heptane, b.p. 98·4°C, and methyl-*cyclo*hexane, b.p. 101°C. The first two, boiling within a fraction of a degree of pure benzene, would be quite impossible to eradicate from benzene by fractionation were it not for the fact that they form azeotropes with benzene, boiling at 77·4° and 75·2°C respectively, and can thus be eliminated in a defronting stage. The other two, of which methyl-*cyclo*hexane is the most likely to occur, are far removed above benzene in boiling-point and should be able to be fractionated out as intermediates between toluene and benzene, but, in fact, behave far from ideally.

Incidentally, Bradley and Middleton suggested that it might be economical to hydrogenate benzene and toluene fractions separately and thus avoid methyl-*cyclo*hexane and its conversion products becoming mixed with the benzene at all. The saving of steam for fractionating out fronts and intermediates would pay for the additional equipment. Has the Author considered the idea of fractionation before hydrogenation before making high-crystal-point material?

I was puzzled by the analysis figures given by the Author for the crude benzole and hydrorefined raffinate in Table 10. These show that, whereas there was no loss of benzene during hydrogenation, there was a loss of about 16½ per cent of the original toluene. I should have expected from this to see a much higher percentage of methyl-*cyclo*hexane and possibly *n*-heptane in the raffinate. Can the Author indicate where else the 16·5 per cent has disappeared to?

Mr. J. L. Sweeten (A.P.V. Company Limited) :—I should like to concentrate my comments on Table 19, which in my view is the key table in the whole paper. I do not think that it should be taken to be final. Very considerable care should be used in assuming that its figures can be adopted for guidance. My interest in this is as a member of the A.P.V. Company, which is prepared to back opinion with performance, and I would draw attention to the yields shown in Table 19. It may be assumed that Table 10 is an analysis of the feedstock on which this detail of performance has been based, and on the basis of the analysis I have recently given certain guarantees for South Wales benzole. The A.P.V. Company is now building plant, and on this basis the yield of products that it has guaranteed to give is approximately 90·7 instead of the 81·7 per cent in the table plus 2·6 per cent of solvents. This makes rather a great difference to the yield figure, and on the basis that I had this is roughly 4*d.*/gal. It is in favour of acid washing to the extent of 4*d.*/gal, which is a major difference.

The next point is that if one looks at the distillation cost a little further down the table, values of 2*d.* and 0·91*d.*/gal are quoted, but we are not able to suggest that those figures be taken as guidance, because we know that the figure for hydrorefining steam is twice as much, as the steam for the acid washing process, and the quantity of water and electricity would be twice that value also. I do not know how the basis was arrived at, but I feel there is a big difference there that ought to be looked at, and I think it is in the region of 0·9*d.*/gal. If the capital costs are examined, we are prepared to do more than suggest that they are wrong. We are prepared to quote costs that will show they are wrong. For a capital cost of approximately £400,000, the capacity we should provide would be 10 mill. gal/a and not 4·8 mill. For a 4·8 mill. gal plant, we would suggest a price of £300,000 and not £400,000, and this would be split in the ratio of £190,000 at 10 per cent depreciation, and £110,000 at 5 per cent depreciation. I am assuming that in both cases steam was available. On this basis, there is a saving in interest and capital cost of 1·77*d.*/gal. If one adds all these up, it comes to something like 6*d.*/gal in favour of acid washing with a modern continuous plant. I suggest, therefore, that when one looks at current capital costs today—and most of us are concerned with those costs and are not interested in the beauty of the process—it will be apparent that the acid washing process is by no means dead, and will in fact show a better return than hydrorefining on the values quoted here. I should add, in fairness to the Author, that I have adopted his figures of capital cost for hydrorefining, and that it might be said that they were on the over generous side, and that, therefore, my capital charges are excessive against hydrorefining. I still feel, however, that the big difference is that modern benzole plant will recover a much bigger proportion of products from feedstock than that shown in Table 10.

Mr. W. H. A. Webb (Rochester):—I have not had time to read the paper, but from everything that has been said it seems that for the first time one has had a properly balanced story. When the distillation plant at Scunthorpe was being designed, I was associated with the A.P.V. Company, Limited, and carried a large part of the responsibility for the plant. I have been rather severely criticized for over designing, and somebody once said : "If only you had designed that plant properly, it would not have cost so much." It strikes me that perhaps the cost of acid washing distillation plant has been slightly exaggerated, so that the bias on one side is perhaps more than it should have been.

The second point is that when the plant was being designed certain methods of distillation were adopted because of the steam pressure available on the site. Had the steam pressure been higher, one would have had an all-continuous process and a lower steam consumption than the one chosen. I think a good job was done with what was available. Steam costs were certainly higher than they would have been if pressure had been better.

I have been interested in the remarks concerning aromatics and non-aromatics. During the Second World War, I was associated with a plant that took its very low-gravity toluene from Scottish crude benzoles, and we made a proper nitration-quality toluene in a particular type of distillation plant. We took out the associated non-aromatics fairly completely. I have no idea how the mechanics of that process would work in so far as the problems that have been discussed are concerned ; but our figures compared with figures in the United States at that time were much better. When we sort out our aromatics and non-aromatics, I think that British processes are probably just as good as some of the American solvent processes.

Mr. B. A. J. Jones read the following contribution from **Mr. H. W. Cremer** (Bromley, Kent) :—" I have watched for some time the work being done by Mr. Lane and his associates. He has placed us all in his debt by his full account of the reasons for their actions in three different situations, and the conclusions after a period of operation. There is only one word of caution I have with regard to the figures given for operating costs. Mr. Lane has been too modest about the way in which the design and engineering of these works was carried out by his own staff, with the assistance of consulting engineers. It is my view that the capital costs of any of these plants would have come out much higher if a single contract had been placed for a turn-key job. I should hesitate in placing a figure on the difference for each gallon of crude processed, but it would be at least ¼*d*. This represents a very considerable sum over the throughputs discussed. Would the Author agree that he has made such a saving by taking the responsibility for overall planning and engineering."

Mr. Jones then said there was one observation he would personally like to make, relating to the time that elapsed between the placing of orders for equipment in Lincolnshire and at Port Talbot and the commissioning of the plant. A careful engineering programme must have been drawn up to to keep to this, and he would like to know whether the planning and control were covered by Mr. Lane's own engineering services.

Mr. N. A. H. Holt (A.P.V. Company, Limited) :—When comparing the costs of acid washing and hydrorefining, I wonder whether Mr. Lane could say how much sulphuric acid he assumes should be used for the acid washing. My reason for asking is that, in the case of the South Wales crude, the sulphur as thiophen content can be seen from Table 10 to be about 0·1 per cent w/w (although, presumably, the actual figure is that of 0·23 per cent as quoted in Section VI (2) (A) and (6) (B), whereas the crude benzole used by the Lincolnshire Chemical Company at Scunthorpe has a sulphur as thiophen content of about 0·5 per cent w/w. Of course, the amount of acid required to give a certain grade of benzene product depends very much on the original sulphur content of the unwashed material and, because of this and other differences in the natures of the two crude benzoles, it is not quite fair to compare hydrorefining practice at Port Talbot with the acid washing practice of the Lincolnshire Chemical Company.

Secondly, could the Author say how much of the ethyl benzene produced during the hydrorefining process can be tolerated in the xylene products? If there is more than a certain amount of ethyl benzene in xylene, the latter will fail on flash-point specification. Furthermore the ratio of styrene (and coumarone) to xylenes in benzoles varies very considerably. Could it be that in some cases the amounts of ethyl benzene produced during hydrorefining would be too much for the specification?

Thirdly, in the Summary of his paper, the Author quotes a yield of 85 per cent of benzene for hydrorefining, which specifically refers to a crude of low initial aliphatic content, and yet, when he gives his figure of 83 per cent for the yield of benzene obtained on acid washing, he specifically refers to a crude *not* of low initial aliphatic content. However, if the benzole used was a normal coke oven crude, such as the Author describes as being used for hydrorefining, then upon prefractionation and continuous acid washing the yield of benzene of crystallizing-point 5·4°C would probably be greater than 95 per cent of the benzene originally present. Perhaps Mr. Lane would comment on this.

Mr. K. H. V. French (Benzole Producers, Limited) :—I should like to comment on the knotty problem of normal heptane. The Author has referred to difficulty in separating it, and there has been some mention of its boiling-point, which is a fair way away from that of benzenes. It is now becoming fairly well established that there is no azeotrope between benzene and heptane, but the shape of the vapour equilibrium curve is such that the relative volatility of this mixture becomes close to 1 as the concentration of heptane becomes low. This being so, it means that minimum reflux for the separation of this mixture becomes high, and it is probable that the Author's plant is not putting on that necessary minimum reflux. In any case, the best he could hope for would be to get hexane into toluene and not into forerunnings as he mentions in the paper.

Sulphur analysis in the region of parts per million, is a big problem and the Benzole Producers, Limited, research paper on the Rhone-Poulenc method was mentioned. We are not satisfied that this method is the best. We should prefer to estimate individual sulphur compounds rather than total sulphur. It has been demonstrated that acid washing or hydrorefining can deal with sulphur compounds, but I still have considerable doubts whether hydrorefining alone will be able to meet the very high crystallizing-point required, and I suggest some ancillary process such as " freezing " will become essential.

Vote of Thanks

Dr. R. A. Mott (President, The Coke Oven Managers' Association) :—In proposing a vote of thanks to Mr. Lane for his excellent paper, I am sure all will agree that on this occasion the high standard of our Joint Technical Meetings has been well maintained.

This problem is a matter of very great commercial importance to the carbonizing industries. Mr. Wilkinson has reminded us that the chemical indstry has, during the last decade, increased its usage of premium-grade benzenes from 10 mill. to 40 mill. gal/a, that anticipated requirements of these pure grades of benzene will in 1965 be 70 mill. gal, and that this will be beyond the ability of the carbonizing industries to provide. Therefore, the chemical industry is clearly going into the production of pure grades of benzene from petroleum.

The comparison between the grades of products that the carbonizing industries produce and those that the petroleum industry can produce is very important. Many will remember that when the chemical industry took up the manufacture of synthetic ammonia it destroyed the value to the carbonizing industries of by-product ammonia. It is important that the carbonizing industries should realize what the position is regarding this important by-product, and should take the necessary steps to ensure that it has a big percentage of the market in the future. It is not our job to make financial decisions, but it is important that we should present the technical information on which financial decisions can be made. I think it will be agreed that Mr. Lane's very full and informative paper gives this sort of information on a very important subject.

We owe Mr. Lane a very great debt of thanks, and therefore I have much pleasure in proposing a hearty vote of thanks to him for his excellent paper, and for the way in which he presented it.

Mr. G. W. Lee (Director, British Coke Research Association) :—It will have been noticed that all the contributors to the discussion, including those whom I might label the " opposition " to Mr. Lane, have emphasized the quality of his paper, and how much they appreciated his presentation of it. We all subscribe to this. I can only add, in support of what the President of The Coke Oven Managers' Association has said, that I have taken some small part in all these Joint Meetings, and I can say quite truthfully that Mr. Lane's paper has been one of the best ever presented to us.

He has that fluidity of command of the English language which, I suppose, is characteristic of tar distillers, and, at the same time, gains the admiration and despair of coke oven managers. We compliment him on his paper and on the very able way in which he has presented it.

I have great pleasure, on behalf of the British Coke Research Association, in seconding the motion proposed by the President of The Coke Oven Managers' Association.

The vote of thanks was carried by acclamation, and the meeting then terminated.

Written Reply

The **Author**, in reply, wrote :—In reply to the many detailed questions, I should like to reiterate the general basis of the preparation of the paper. The figures put forward relate to actual works production figures based on the first six months' operation

of the plant. The analyses put forward are based on G.L.C. analysis, and again the accuracy is only as accurate as the technique employed. It is, therefore, natural that in such a complex mixture as crude benzole there will be certain discrepancies ; I feel these must be accepted.

The major point, which I will take first, is the apparent reluctance to accept the difference of an 8 per cent greater yield of water whites by hydrorefining than by acid washing techniques. Although I find it somewhat difficult to be at variance with such an acknowledged expert as Mr. Claxton, it is not unnatural I should be at variance with the manufacturer of acid washing equipment.

In answering this point, I should like it to be borne in mind that the figures I am referring to are overall works production, from material coming into the gate to going out of the gate, and not an arithmetical formula based on performance of individual units of plant. To illustrate the point, I have shown the overall loss of products by hydrorefining as 3·36 per cent, compared with 6·0 per cent on an acid washing plant. Thus, physical loss accounts for 2·74 per cent of the difference.

In looking at the make-up of the Port Talbot loss of 3·36 per cent, on works production statistics, the loss in the hydrorefining section was 1·07 per cent, the distillation loss was 2·02 per cent and the unaccounted-for loss (i.e., storage loss or discrepancy in the above works figures) 0·27 per cent. I think these figures are creditable for the first six months' working, but there are obvious reasons to believe that the distillation loss and unaccounted-for loss will be reduced when the management and plant operators get more familiar with their problems. One can see that, ultimately, the overall works loss will settle down at 0·5 per cent lower, which would compare with the overall works loss at Cadishead, which is regularly 2·0 per cent or lower. This would mean that acid washing would have to run on an overall loss of 5·5 per cent to redress this difference, and these would be considered good figures on any standard. In addition, there are 1·5 per cent of forerunnings, so that these two together add up to more than half the difference in yield between the two processes, and we are looking for 3·8 per cent for other reasons. In addition, the crude contains 1·33 per cent of styrene, and this would, in fact, by acid washing, account for at least 2 per cent of the crude as loss. This would allow a 1·8 per cent difference for complex formation in the acid washing technique and for sulphonation. Little mention is made by the staunch defenders of acid washing of the sulphonation, and, despite our experience with acid washing at Scunthorpe, we, too, are a little unsure on this. For the pulsed washing column, I was given the figure of 3 per cent, although I should like to believe that this could be a little high.

Personally, therefore, I have considerable confidence in the figure of 8 per cent difference put forward, although it must come as a considerable shock to some people who relate it only to individual pieces of equipment and do not look upon the whole picture of the overall works yield.

A second point I should like to make is that we have been discussing these costs making No. 2 pure benzene. We all realize that in future there will be much more stringent specifications for benzene than the 100 p.p.m. of sulphur as thiophen mentioned in the No. 2 specification. Hydrorefining does start at the figure quoted, at 1·5 p.p.m., and with acid washing there is most probably considerable extra cost and loss of yield in getting a similar benzole to the one made at Port Talbot.

Coming now to the more detailed questions, in reply to Mr. Illingworth, the Cadishead hydrorefined raffinate contains about 25 to 30 per cent of paraffins that are essentially straight-chain paraffins. The normal petroleum industry practice for aromatizing its feedstock is to choose a high naphthenic feed, although the same techniques can be applied with less effect to straight-chain paraffins. I am a little doubtful what proportion of the paraffins would be converted to aromatics, but I am of the opinion that it would be low. However, I have been reassured by Dr. McNeil of the Coal Tar Research Association, that, although his work is as yet in its infancy, he is hoping that catalysts and conditions can be arranged to give a 30 per cent conversion of the paraffins present, which would give more than enough hydrogen as a by-product to run the hydrorefining plant, and the value of the hydrogen in this might well be the major item rather than the one-third aromatization of the paraffins present.

Water injection for cooling is a chemical engineering feature incorporated by the contractor, and accounts for a further question that was asked on what is the condensate in the cost table. This is, in fact, a pure water for injection for cooling. In addition to cooling, it is also a convenient way of getting rid of the ammonia formed in the hydrogenation of the pyridines present.

Mr. Pater rightly draws attention to the after treatment (i.e., the distillation of the hydrorefined product) in the case of crude benzole of high paraffinic content. At present, we have found no formula for making pure products from our Cadishead 30 per cent paraffinic material. It is my belief that, even employing azeotropic techniques, as suggested by Mr. Webb, the problem would be difficult, but not impossible, and would be purely an economic consideration. It is my belief that the technique of solvent extraction, given suitable solvents, is much more adaptable for this process. I think, as Mr. Ryder points out, one must closely watch whether the overall economics of selling paraffinic material and aromatic material are in fact less attractive than selling the mixture at a discount. On the ratio of price as between coke oven gas and hydrogen, I suggest that the price of hydrogen would have to be approximately 4*s.*/1,000 ft³ to show similar economic advantages over this process, although a lower figure of 2*s.*/1,000 ft³ might also be substantiated if one were to take account only the hydrogen used for the hydrogenation of the crude benzole.

I thank Mr. Pater also for drawing my attention to the discrepancy, which has now been corrected.

Mr. Victory is quite right in his assumption that, to minimize the loss of benzene in the tail gas, the pressure is let down from the maximum possible pressure. The pressure is higher at Port Talbot because, when using coke oven gas, there is a large quantity of gas as tail gas, whereas at Cadishead the tail gas is only the gaseous hydrogenation product and is less than 0·1 of the Port Talbot quantity. The problem is, therefore, very much less at Cadishead and there are operational advantages by working at the lower pressure when removing the hydrogen sulphide by distillation. The fact that we have two let-down stages at Port Talbot, as compared with Cadishead, is also connected with the use of coke oven gas rather than pure hydrogen : thus, one of the stages is saved at Cadishead.

With regard to the pretreatment, as this was a late addition on the plant I have no real idea of its exact function. However, it obviously does useful work in cleaning up the vaporized benzole and coke oven gas mixture, and we have recently incorporated such equipment on the Cadishead plant.

I would agree that, academically, there could be some advantage in defronting the crude benzole before hydrorefining, but, in view of the very low cost of hydrogen as coke oven gas and the fact that the reaction is providing heat, I believe the cost of hydrogen in coke oven gas and the additional heat is lower than the cost of predistillation, which would then also leave one with a forerunnings fraction, which is always a disposal embarrassment when practising defronting. This point is also brought out by Mr. Townsend, who suggests the recovery of di*cyclo*pentadiene from the fronts. Unfortunately, this recovery involving carbon disulphide removal as well, is not a very pleasant process and the best known process is patented to Y.T.D. I believe, therefore, that on economic factors alone, it is better and cheaper to hydrogenate rather than defront, with the factor that the less the number of operations involved, the less control, manpower, handling, *etc.*, are necessary with their associated loss of product.

Mr. Wilkinson asks for the individual content of benzene, toluene and xylene in the Port Talbot crude. A typical crude is shown in Table 1 for Cadishead, and in Table 10 for Port Talbot.

Mr. Ryder asks whether I am referring to the sulphur content of the raffinate or in the benzene fraction. In Section V(8), it does, in fact, refer to the sulphur content of the total raffinate, but, as the sulphur is generally believed to be totally thiophen when getting down to these low limits, the sulphur contents of the raffinate and the pure benzene are always approximately equal, although by distillation techniques, by fixing the right temperatures for distillation, it is possible to minimize and, to a certain extent, control the amount of thiophen going into the pure benzene.

The cracked petroleum, to which Mr. Townsend refers, had a very high bromine number, which was essentially associated with olefins and, particularly, di*cyclo*pentadiene. This gave considerable trouble, but we had considered techniques of minimizing and of overcoming it, but the contract for the raw material finished before these were put in hand.

To the best of my knowledge, carbon monoxide has no harmful side effects other than as a user of hydrogen, unless it is in such high proportion as to influence Boudouards reaction for coke formation, but this would mean that to maintain 18 to 21 atm partial pressure of hydrogen, large volumes of gas might of necessity have to be used.

Regarding the cleaning of the reactor, we believe the problem at Cadishead was associated with the cracked petroleum feed, and this required cleaning quite substantially every three to six months. However, I would say that at Port Talbot, after nine months' running, there was only a small powdery deposit on the top of the reactor and there had been little degradation of the catalyst, except for a little breaking-up in the first few inches. At Cadishead, we did make alterations at one time to introduce the hydrogen as suggested before the evaporator. It is quite right that, theoretically, this should be a better place for injection. However, we found we tended to use more hydrogen, although this was never clearly explained; therefore, we reverted to using it between two stages of the reactor. The six-stage evaporation has no special reason. The evaporator employed by Lurgi is most ingeniously designed. Basically, the six stages are employed as part of this technique and to keep the size of individual vessels down on the plant.

My figure of 10 mill. gal/a or more throughput is based on the usual contention of capital and running-cost considerations for large-sized plants. For instance, at Port Talbot, the works variable costs related to throughput are only 20 per cent of the overall expenditure, and this, I believe, would show the advantages of the larger plant. I believe also that in a very large plant being built in Czechoslovakia by Lurgi the larger size has made it possible to utilize the heat of reaction alone and preclude the use of additional heat to make up radiation loss.

In reply to Mr. Bradley, we all know the useful work and number of papers given by Mr. Bradley, and it is not unnatural that we have a good deal of agreement. However, I am not impressed by the contention that an acid wash before or after hydrorefining and working at a low pressure is a particularly good idea. This seems neither to give the best conditions of hydrorefining nor the best conditions of acid washing, but to give the problems associated with each. There is, firstly, the disposal of acid effluent, and, particularly if washed before hydrorefining of naphthas, there would be some polymerization, losing the benefit of yield. It would also mean that there could be some risk of the material being acidic and that the distillation plant would have to be made of more costly materials of construction. If there were to be a saving of hydrogen, then the product would be incompletely hydrogenated and would not hold the colour stability of the fully hydrogenated material. I cannot feel that it would substantially decrease the capital cost as the plant would most probably be designed to withstand 50 atm in any case.

The reason interest on capital is omitted is that, with the constant changing of the bank rate, I believe that this is outmoded as a form of presentation and as a cost item. What is important, however, is the return on capital invested, which will include any interest necessary on the investment. This figure should obviously be not less than 15 per cent yield on capital invested, and a more normal figure used in the chemical industry is 25 per cent. However, in showing costs one should show that interest on capital is not considered a working cost, as has been done in this paper.

Mr. Townsend asks why we do not have an individual hydrogen plant to feed the hydrorefining plant. We did look into this, but, bearing in mind my previous comment on return on capital invested, we felt that, although hydrogen could be cheaper, it would not give the requisite 25 per cent on capital; therefore, we preferred not to buy a hydrogen plant. We are also hopeful that in the future we shall be able to get a source of hydrogen-rich gas either as a by-product of one of our own processes or from coke ovens or town gas supply. The present town gas at Cadishead has only a 35 per cent hydrogen content, and then there are the natural difficulties of disposing of the tail gas. This could have been a difficult chemical engineering project.

With regard to catalyst regeneration, with the new catalyst we are told this is approximately every three years. However, due to taking the cracked petroleum feed at Cadishead, the regeneration has been done at more frequent intervals and this has averaged at about every nine months. The job is quite easy and there are definite signs in the working of the plant when regeneration is necessary by the falling off of the quality of the products. It is true that one would expect corrosion in the reactors, and for this reason the Lurgi Company has used special materials of construction, which are among the few things on the secret-list so far as we are concerned. After two years' running, there is no obvious evidence of any sort of corrosion taking place.

With regard to the removal of naphthalene from the feedstock, when we are down to limits of less than 2 per cent it is extremely difficult to improve on this by refrigeration and it is likely to be more expensive than hydrogenating it, although we appreciate that this is an undesirable side effect. Little actual tonnage of naphthalene is lost in this way and, although we do not like it happening, we are constantly trying to improve our prefractionation to get the percentage of naphthalene in the feedstock lower.

With regard to the alkali wash for removing last traces of hydrogen sulphide, the early Lurgi plants, including that at Cadishead, had a system for a caustic and water wash of the raffinate. However, their chemical engineering and confidence had increased so much regarding the efficiency of the hydrogen sulphide removal section, that no caustic washing vessels were installed at Port Talbot. It is not an unknown occurrence for hydrogen sulphide to come over in the raffinate, but, because of the trouble of recycling contaminated raffinate, it is happening less often and is due to a really bad operating error.

I think that in my opening general remarks, I have dealt with Mr. Claxton's criticism, and he has kindly answered the question regarding the de-methylation process developed in America.

Although Mr. Claxton seems now to lean towards acid washing, where previously his advice was for hydrorefining, I do not believe that there has been enough work done yet on the pulsed columns to show a comparison on the basis shown in the paper, which shows the overall working losses and costs and is not related to an arithmetical progression of the various steps involved. However, I would agree that the position should be looked at carefully by each operator and it will be noticed that even in our own reasoning we had an acid washing plant sandwiched between two hydrorefining plants. This is a decision that we do not regret; we know in retrospect that it was a right decision for the case in question, as was our decision for the hydrorefining plants.

I am afraid, however, I must still maintain that there is a loss of 8 per cent of water whites when one looks at overall works products, but whether this in itself justifies the extra capital cost and slightly more difficult after-treatment is a decision each operator must make. However, I hope the paper gives sufficient information to anybody contemplating such a decision between hydrorefining and acid washing for him to see what actually happens as works practice.

Mr. Ritson again brings forward the question of pre-fractionation. It is my personal opinion, and, in answer to several previous questions, I do not think that there is any worth while reason for doing this on economic factors alone. However, I am against complication in any form. Each extra operation means more jobs to control and more labour. There is also a source of further loss in handling and pumping, and one of the beauties of hydrorefining, as pointed out by Mr. Scott, is that there are only two operations, namely, hydrorefining and distillation, compared with three in acid washing by modern methods, i.e., pre-fractionation, washing of individual products, usually at least three different streams, and blow-over and final distillation. The storage and overall statistical control are, therefore, made more simply in the hydrorefining process.

I cannot possibly blame Mr. Sweeten for the fine piece of salesmanship he has produced, and, although his company constructed a wonderful plant for us at Lincolnshire Chemical Company, Limited, with which we are delighted, and know that his company has a very high reputation in the benzole refining industry, I cannot possibly agree with his figures or logic. My paper, as I have said before, is based on actual works production; it is not based on figures based on a guarantee of a plant not yet constructed for an unspecified crude benzole. I have no idea how it compares with that being refined at Port Talbot. In my opening remarks, I have covered many of the points Mr. Sweeten brings up, but I am afraid that, having built both hydrorefining and acid washing plants, I am firmly entrenched in the 8 per cent difference in water white products between hydrorefining and acid washing when based on the crude in question and on an overall works yield, irrespective of what arguments might be advanced over individual sections.

I should say that, so far as capital cost is concerned, before making the decision for hydrorefining at Port Talbot, we did work

out a full scheme for acid washing, and the quotation came from Mr. Sweeten's company. I am therefore, basing this on actual facts, and I am interested to hear that prices have come down, at least in theory.

Also, I cannot agree that the distillation costs are as vastly dissimilar as he suggests. In fact, now that we are beginning to understand the distillation section better, there is virtually no difference in the cost of distilling the hydrorefined material and the acid washed one, and one must remember that there is only one distillation not two. The ingenuity of the design at Port Talbot is that there is only one reflux used both for the complex fore-runnings section and for the main toluene—benzole split. With the forerunnings being 4 per cent of the crude, and the benzene fraction in excess of 60 per cent of the crude, the reflux ratio on the forerunnings section is 15 times greater than on the benzene cut, and this ingenious design does mean in fact that one obtains a very substantial reflux ratio on the forerunnings for quite a modest reflux ratio for the benzene—toluene split. However, I have no axe to grind, for in most cases for small operators the choice would be naturally towards acid washing, and this would be, with the development done on acid washing and particularly by Mr. Sweeten's company, a very sound choice. However, I am equally certain that should anybody, particularly with larger throughputs, install a hydrorefining plant, they would find an equally sound and most probably slightly more profitable proposition.

Mr. Webb, who I have known for a good many years, but not, unfortunately, since he has been a whisky refiner, has done a lot of work on azeotropic distillation and we know the good work he did in up-grading low-gravity toluene when it was so short between 1939 and 1945. I am certain that in this field he could offer a design competitive with anything in the World. However, I have some doubt as to whether azeotropic distillation is the right approach for highly paraffinic crudes, and feel that solvent extraction, if the many problems can be overcome, is a more desirable chemical engineering project.

I am sorry that neither Mr. Cremer nor Mr. Warner could be present and it is most probably remiss of me not to mention in my paper that they were consulting engineers both for our Lincolnshire Chemical Company development and the Port Talbot plant, although at Cadishead we did use our own engineering staff. It was fairly obvious, with an engineering staff of only two at Lincolnshire Chemical Company, that a project of this size would be beyond them, and we certainly found it quite a strain even with a larger staff at Cadishead. It is obviously less expensive if one has drawing-office and engineering staff to carry out the project oneself, and, in fact, one gets it done in the manner that one requires. However, there is no doubt that the use of consultants at Port Talbot greatly assisted us in the speed with which the plant was constructed, and we were able to incorporate not only our own idea in design and specifications, but to have the benefit of a wider field of experience brought in by the consultants. The bulk of the expenditure was on plant, and both Lurgi and Dempsters quoted for a turn-key plant, and the engineering of these projects was left to them. It was, therefore, only the civil engineering and services and co-ordination that was handled by the consultants, although, particularly at Port Talbot, the speed of construction and the ease with which it went was due to fine team work amongst the various site contractors, consultants and ourselves. With regard to cost, I would certainly say that the way we dealt with the project saved the consultants' fee in capital cost, but, whether it went further than this as compared with giving it to one main contractor, I should not like to say. However, it did save a considerable amount of work and, being of an independent nature, we never like to be entirely in the hands of one contractor for a whole project.

In answer to Mr. Holton, the figures for acid washing costs were based on the operation of acid washing Steel Company of Wales' crude at Lincolnshire Chemical Company, Limited, and, although different products were made, I did not in fact break it down into specific costs, but used an actual works cost.

Although we had some fear that the ethyl benzene in the xylene would affect the flash-point, this has not proved to be the case; in fact, it is interesting to note that the ethyl benzene content of our xylene is lower than that of two samples of petroleum xylene we have had for examination.

I cannot, however, agree about Mr. Holton's figure of 95 per cent of nitration-grade product by acid washing when related to the overall works products. Needless to say, the 85 per cent he quoted it against was in making a 5·4°C crystallizing-point benzene.

Proceedings of Council

The following matters of general interest were among those dealt with by the Council:—

COUNCIL MEETING 31st OCTOBER, 1961

Reports were received of meetings of the following Committees of the Institution :—

(1) GAS DISTRIBUTION COMMITTEE, on 13th October, 1961

Mr. G. G. Warne (Southampton) had been appointed to the Committee as the representative of the Southern Gas Board. Matters currently being studied by the Committee included the record of broken mains, mains and services surveys, and the safe operation of district governors.

(2) GAS EDUCATION COMMITTEE on 17th October, 1961

Success in a sandwich course in Gas Engineering (Utilization) of four years' duration, conducted by the Westminster Technical College, was approved as a suitable examination qualification for Associate Membership of the Institution.

Reports were received of arrangements that are being made for a short refresher course for gas utilization engineers, to be held at Pembroke College, Oxford, from 8th to 14th April, 1962. It is hoped that a short course for teachers in gas technology and gas fitting will be held at Norwich in July, 1962.

A report was received of a meeting, on 20th October, 1961, of the Papers Committee of the Joint Committee of The Institute of Fuel and The Institution of Gas Engineers for the administration of a joint conference on gasification to be held in September, 1962.

Two representatives, Mr. G. E. Currier, O.B.E. (*Honorary Secretary*) and Mr. N. C. Sturrock (Chairman, Manchester and District Section) were appointed as the Institution's delegates to attend the Health Congress of the Royal Society of Health at Scarborough on 9th to 13th April, 1962.

A report was given of a meeting of the Chairmen and Secretaries of the standing committees of the International Gas Union, held in Brussels on 27th October, 1961.

COUNCIL MEETING 21st NOVEMBER, 1961

A short meeting of the Council was held at Church House, Westminster, London, S.W.1, immediately preceding the opening session of the 27th Autumn Research Meeting to consider matters relating to the programme of the Meeting.

27th Autumn Research Meeting

The 27th Autumn Research Meeting of The Institution of Gas Engineers was held in Church House, Westminster, London, S.W.1, on Tuesday and Wednesday, 21st and 22nd November, 1961, under the presidency of Thomas Charles Battersby, Esq., M.B.E., M.I.Mech.E., M.I.Gas E., Controller, Tottenham and Watford Divisions, Eastern Gas Board, who occupied the chair at those sessions that were held in the Assembly Hall.

Albert Edward Haffner, Esq., Ph.D., B.Sc., Senior Vice-President of the Institution, occupied the chair at parallel sessions held in the Hoare Memorial Hall of Church House.

At the 27th Research Luncheon, held at the Savoy Hotel, Victoria Embankment, London, on 21st November, 1961, the President of the Institution took the Chair and proposed the Loyal toast and that to " The Guests ". On behalf of the guests, the response was made by Sir Harold Roxbee Cox, D.Sc., Ph.D., D.I.C., M.I.Mech.E., F.R.Ae.S., F.Inst.F., F.I.Ae.S., Chairman of the Research Council, Department of Scientific and Industrial Research.

An informal toast to the President was proposed by Sir Henry Jones, M.B.E., M.A., Chairman of The Gas Council, which Mr. Battersby briefly acknowledged.

Detailed reports of the proceedings at the business sessions and the Luncheon will appear in the Institution *Journal* for January, 1962.

The reports and papers presented at the Meeting, together with appropriate discussions and written replies thereto, will be reproduced in the January, 1962, and subsequent issues of the Institution *Journal*.

Membership Subscriptions

Owing to the substantial increases in postal charges, an attempt is being made to economize by arranging that, as from 1st January, 1962, individual subscription applications will not be forwarded to each member as heretofore.

A printed circular will be enclosed with the January issue of the *Journal*, reminding members of the annual subscription due for each class of member. Members who have provided for the payment of their membership subscription by means of standing instructions to their bankers should ignore this circular.

An official receipt in acknowledgment of remittances by cheque will be issued only upon request.

Personal Notes

Appointments

(a) By The Gas Council :

London Research Station—
Assistant Director : **W. B. S. Newling, M.A.**

(b) By Area Gas Boards :

NORTH EASTERN GAS BOARD
Bankside Works, Hull—
Works Manager : **D. K. Blears.**

EAST MIDLANDS GAS BOARD
Southern Production Region—
Assistant to Production Engineer : **J. Parry, B.Sc.**
Notts. and Derby. Division
Chesterfield District—
District Distribution Engineer : **R. L. Carter.**

WEST MIDLANDS GAS BOARD
Director of Marketing : **W. H. Elkerton.**
Foleshill Works, Coventry—
Assistant Engineer : **R. H. R. Tomkins.**

WALES GAS BOARD
Cardiff Works—
Senior Technical Assistant : **V. L. Owen, B.Sc.**

NORTH THAMES GAS BOARD
Group Engineer : **D. C. Bertram, G.M., B.Sc.**
Southall Works—
Station Engineer : **T. Bispham, B.Sc.**
Deputy Station Engineer : **G. F. I. Roberts, B.Sc.**
Kensal Green Works—
Station Engineer : **C. P. N. Cox.**
Senior Assistant Engineer : **T. C. Tweddell.**
Nine Elms Works—
Station Engineer : **A. F. Grant, B.Sc.(Eng.)**
(with the status of Group Engineer)
Brentford Works—
Station Engineer : **R. T. de Gruchy, B.Sc.**
Slough Works—
Station Engineer : **G. L. Braidwood.**
Stratford Works—
Station Engineer : **M. Gregory.**
Bromley Works—
Deputy Station Engineer : **R. Bairstow, T.D.**
Bow Common Works—
Assistant Engineer : **P. C. Egan, B.Sc.**

SOUTHERN GAS BOARD
Commercial Manager : **B. G. H. Clegg, M.A.**
Service Manager : **D. M. Foulis.**
Industrial and Technical Sales Officer :
W. A. Lloyd-Dodd, B.Sc.
Distribution Engineer : **G. G. Warne.**
Assistant Distribution Engineer : **J. Gadd.**
Wight, Portsmouth and Southampton Regions—
Regional Distribution Engineer : **D. J. Butler.**
Dorset and Bournemouth Regions—
Sales Manager : **J. F. Doran.**
Poole Works—
Station Engineer : **R. L. Hall.**

(c) General :

Mr. W. A. Campbell (M) has been appointed Deputy City Gas Engineer and Manager with the Belfast Corporation Gas Department, Northern Ireland.

Mr. L. Clutterbuck (AM) has taken up an appointment as Works Manager with Bitmac, Limited.

Dr. W. T. Cosby (M) has been appointed to the board of directors of W. C. Holmes and Company, Limited.

Dr. D. G. Edwards, B.Sc. (AM) has taken up an appointment as Abstractor with the British Coke Research Association.

Mr. J. B. Friggens (AM) has been appointed as London Manager of W. J. Jenkins and Company, Limited.

Mr. C. A. Lewis (A) has taken up an appointment as Technical Consultant with G. and R. Gilbert (Industrial), Limited, Hackbridge.

Mr. R. Miller (AM) has been appointed to the position of Manager of the Chemical Division of W. J. Jenkins and Company, Limited.

Mr. A. J. Penston, M.A. (A) has been appointed to succeed Mr. M. W. Burt, M.B.E., as Director and Secretary of the Society of British Gas Industries.

Mr. B. I. Stone (A) has been appointed Director of The Bryan Donkin Company, Limited.

Retirements

Mr. F. B. Holmes (M) has retired from his position as Managing Director of W. C. Holmes and Company, Limited. Mr. Holmes continues to be Chairman of the company, of the holding company, B.H.D. Engineers, Limited, and of those member companies of which he is at present Chairman.

Mr. W. M. Paterson (M) has retired from his position as Distribution Engineer with the Sussex Division of the South Eastern Gas Board.

Mr. A. Robin (M) has retired from his position as Deputy City Gas Engineer and Manager of the Belfast Corporation Gas Department, Northern Ireland.

Addresses Required

The Council would be grateful to receive the present addresses of the following members, last heard of in the places given in brackets :

Mr. W. G. Bignell (Eastbourne).
Mr. E. Clark (Bridlington).
Mr. S. W. Hayes (Braunton).
Mr. D. Helliwell (Halifax).
Mr. P. W. Hugh (Middlesbrough).
Mr. C. Kay (Cambridge).
Mr. N. P. Shepherd (London).
Mr. A. M. Steel (Germany).
Mr. L. H. Watson (Coventry).

Institution Register

Transfers

(a) To Membership :

Agrawal, Suraj Narayan, B.Sc., (Bombay, India)
Buckley, James Arthur (Woodford)
Cox, Charles Peter Northwood (London)
Dewhirst, George Milner (Aylesbury)
Fairman, John Robert (Southampton)
Ireland, James Lawrence (Lincoln)
Moorley, Thomas Edward (Oxford)
Moys, Herbert William, B.Sc.(Eng.) (London)
Patching, Harold John (Montevideo, Uruguay)
Rimmer, George James (Hinckley)
Shipley, Thomas William Edgar (Keighley)
Wall, Charles Denis (Maidstone)
Whitelock, Michael James Simpson, B.Sc. (Wetherby)

(b) To Associate Membership :

Ahmad, M. Bashiruddin, B.Sc., B.E. (Karachi, Pakistan)
Haken, Arthur Maurice (Crowborough)
Rowe, Norman (Stoke-on-Trent)
Stevens, Michael Henry (Colombo, Ceylon)

Admissions to Studentship

Duckworth, Barry (Accrington)
Gray, Gordon David Victor (Bournemouth)
Griffiths, John Granville (Flint)
Matthews, Kenneth (Swansea)
Tresilian, Anthony George Brogan (Cardiff)

Deaths

The Council regrets to announce the death of the following members of the Institution :

(*a*) *Honorary Member:*

Andrews, Roland Stuart, C.M.G., D.Sc. (Melbourne, Australia) HM/1960 ; died 14th October, 1961.

(*b*) *Member:*

Elliott, Allan William (Derby) M.1943; died 3rd September, 1961.

(*c*) *Associate Members:*

Chisholm, William Wallace (Langley) AM.1924 ; died 28th July, 1961.

Cunnold, Clive Clement (London) AM.1928 ; died 30th July, 1961.

Pickthall, Tom Watson (Bury) AM.1938 ; died 25th September, 1961.

(*d*) *Associate:*

Thorne, Reginald Albert William (Storrington) A.1951 ; died 6th July, 1961.

Arrangements for Binding the Journal

1961 Full Binding Service

In accordance with the previous notice in the *Journal*, arrangements have been made with Straker Brothers, Limited, 194 to 200, Bishopsgate, London, E.C.2, for binding the 12 monthly issues for 1961, in an attractive blue buckram similar in appearance to that traditionally used for the bound volumes of the *Transactions* of The Institution of Gas Engineers. The binding case will be embossed in gold with the crest of the Institution and an appropriate title. Indexes are included in this issue, and a title page will be supplied by the binders, who will also remove the advertising matter and paper covers from each issue.

Members and subscribers who wish to avail themselves of this arrangement should complete the top portion of the yellow order form enclosed with this issue and send this to Straker Brothers, Limited, with the 12 issues for 1961, and a remittance of £1 8*s*. 0*d*., which includes packing and postage of the bound volume. Sets for binding should not be sent to the offices of the Institution.

1961 Binding Case Only

For readers who wish to make their own arrangements for binding the *Journal*, the binding cases described above are also available separately from Straker Brothers, Limited, at a cost of 8*s*. 0*d*. each. A title page will be supplied with each one. Those adopting this procedure should indicate their requirements under section (*b*) of the order form enclosed with this issue, and return it to Straker Brothers, Limited.

1962

In 1962, a third method of obtaining a bound annual volume of the *Journal* will be available. The services described above will also be offered at the end of 1962, but, in addition, members and subscribers who order **now** may reserve a bound copy of the *Journal* in addition to their monthly issues (which will be posted as usual). The material for this bound volume will be laid aside by the printer each month until December, 1962. Readers who adopt this alternative will then receive a bound copy in perfect condition, and their monthly issues will remain continuously available. They will also be saved the trouble of posting copies back to the printer for binding.

The charge for a bound volume prepared in this way will be £1 13*s*. 0*d*., but **it must be emphasized that this service is available only to members and subscribers to the monthly issues.** Bound volumes will not be sold separately, and will only be reserved for readers who place a firm order **now.** Those who wish to take advantage of this service, should complete the lower portion of the yellow order form enclosed with this issue and return it to the Secretary, The Institution of Gas Engineers, 17, Grosvenor Crescent, London, S.W.1, unless they have already indicated that they require this extra bound volume. In order that sufficient additional copies of the January, 1962, and subsequent issues may be printed, it is essential that these orders should be received by the Secretary **not later than 15th December, 1961.** Invoices will be rendered to those who use this method of ordering bound volumes, so that no money need be sent with the order.

News from the Gas Industry Overseas

International Gas Union

The Council of the International Gas Union met in Bucharest on 3rd October, 1961, at the invitation of the Association Scientifique des Ingénieurs et Techniciens de la République Populaire Roumaine.

The meeting was held under the chairmanship of Ir. J. van Dam van Isselt (The Netherlands), President of the International Gas Union, with Monsieur R. H. Touwaide (Belgium), Secretary-General, in attendance. The delegates of the national gas associations of Czechoslovakia, France, Great Britain, The Netherlands, Poland, Roumania and the Union of Soviet Socialist Republics were present.

Ir. I. Niculescu, President of the Petroleum and Gas Section of the Association Scientifique des Ingénieurs et Techniciens de la République Populaire Roumaine, welcomed the members of the Council and expressed his pleasure at receiving them in Bucharest.

The Council examined the constitution and membership of the standing technical committees set up at the meeting in London, in March, 1961, and decided that they should commence their work as soon as possible. The President, Ir. van Dam van Isselt, convened a meeting of the Presidents and Secretaries of the standing committees in Brussels on 27th October, 1961, for this purpose.

The President and members of the Council expressed their congratulations to their Swedish colleagues on the success of the Stockholm Conference, held in June, 1961. A useful exchange of views took place concerning lessons to be drawn from this Conference, and various aspects of the organization of the 9th Conference, to be held in Amsterdam in June, 1964, were considered.

The next meeting of the Council will be held on 28th March, 1962, in the Federal Republic of Germany.

Abstracts of Overseas Papers

The full text of articles abstracted below may be consulted in the Library of the Institution.

Ample Future Gas Supply Assured.

A.G.A. Bureau of Statistics. *Amer. Gas Assoc. Monthly*, 1961, October, 21-22, 30.

The following figures are quoted for natural gas production in the United States of America :—

Year	Trillion ft³
1945	4·8
1955	10·2
1960	13·1
1970 (estimated)	18·5
1980 ,,	24·0

The proportion of the total energy demand accounted for by natural gas is given as :—

Year	Per cent
1950	17·5
1960	28·3
1970 (estimated)	33·1
1980 ,,	34·1

Proved recoverable reserves of natural gas at the beginning of 1961 are estimated at 263 trillion ft³, but the ultimate reserves, based on estimated ultimate reserves of crude oil and assuming a gas : oil ratio of 7,000 ft³/barrel, are estimated at 1,113 trillion ft³, making a total remaining U.S. reserve of 1,376 trillion ft³. It is anticipated that, by 1990, some of the natural gas will be supplemented by synthetic gas derived from coal, oil, shale, *etc.* This quantity will be small at first, but will increase rapidly and, by the year 2,000, may represent as much as 42 per cent of the total gas supply. These figures do not include significant reserves available in Canada, where ultimate reserves may be of the order of 300 to 1,000 trillion ft³.

Technical Problems in Town Gas Production from Petroleum Products.

H. Deringer. *Monatsbulletin*, 1961, October, 235-243.

After indicating some of the advantages of petroleum products over coal for town gas production (easier handling and smaller quantity of raw material, independence of coke market, reduced operating time), the Author poses and discusses a variety of questions affecting the operation of the alternative processes available, primarily from the point of view of the Swiss gas industry. Duplicate installations, working together in parallel (for an 8-h shift), are preferred to a single installation working a 16-h shift. Difficulties due to the variable composition of refinery gases are mentioned, and more detailed consideration given to the choice between light spirits and fuel oils as raw material. Light spirit can be gasified in cyclic or continuous plants, whereas fuel oils can only be gasified in cyclic plants. Important characteristics of the suitability of a raw material are the elementary analysis (particularly carbon : hydrogen ratio and sulphur content) and boiling-range. The selection of a light spirit for cold carburetting requires a knowledge of its dew-point in order to avoid condensation either at atmospheric pressure or at higher storage pressures. Warming the gas to prevent condensation is discussed. The Author discusses also the thermal principles underlying the gasification of light spirits and fuel oils and the relative advantages of the continuous and cyclic processes. He refers also to the causes and prevention of cooling or overheating of the catalyst and to the operating principles of a new continuous plant, now under construction, for daily production from light oil of 3½ mill. ft³ of town gas of about 500 Btu/ft³.

Notes on the Teletransmission Equipment of the Network Carrying Gas from Lacq.

J. Penaud. *J. des Industries du Gaz*, 1961, October, 423-438.

The network carrying gas from Lacq across France includes the production centre at Lacq, underground storage at Lussagnet, a variety of compression stations, metering-points and the terminals of the main pipelines and branch lines. The purpose of teletransmissions is to convey to control centres the various information vital to the administration of the network and to carry from these centres appropriate instructions and even to control automatically certain operations. The more usual means of teletransmission are radio (with a permitted frequency of 80 Mc/s and a range of about 30 km), or metallic conductors (either using existing Post Office lines or laying private lines) ; in regard to the latter, the Author discusses the relative advantages of above-ground and buried cables, number and diameter of strands, frequencies, and simultaneous telephony and telemeasurement on a single circuit. Technical details are given of systems of telephony, telemeasurement and telecontrol, with explanatory diagrams. Finally, the Author outlines, with drawings, the types of buildings suitable for housing the teletransmission apparatus.

Supplying Safe Gas from the Viewpoint of Gas Undertakings.

W. Attig. *Gas- u. -Wasserfach*, 1961, 20th October, 1116-1122.

In the Federal German Republic, in 1959, 48·5 per cent of fatalities were due to road accidents, 28 per cent to falls, 1·1 per cent to town gas, and 0·9 per cent to electricity. Disproportionate publicity is given in the press to gas fatalities, which have actually fallen from 35·7/milliard m³ in 1950 to 22·6/milliard m³ in 1958. Of the gas fatalities in 1958, 67 per cent were in the age group from 60 upwards, partly owing to declining faculties, but partly also due to unfamiliarity with modern devices and to imperfect appliances found in the homes of the aged. The incidence of fatalities decreases disproportionately with a reduction of the carbon monoxide content of the gas, but almost complete removal is costly, particularly for a country like Germany, where about 80 per cent of gas is used in industry. The Author discusses the report of the I.G.U. Committee on Safety Conditions, agreeing with the recommendation to reduce carbon monoxide content to 10 per cent and pointing out that increasing use of natural gas, petroleum enrichers and coke oven gas, and the introduction of carbon monoxide conversion in gasification processes will help in this direction. In Germany, however, particular importance is attached to ensuring the completely safe operation of domestic appliances, and the German Gas Association is working actively to this end.

Completely Safe Operation of Domestic Gas Appliances.

W. Stehn. *Gas-u. -Wasserfach*, 1961, 20th October, 1123-1129.

German DIN Standards already require the incorporation of safety ignition devices for gas water heaters, space heaters, washing machines and the larger sizes of refrigerator. The gas cooker presents more difficulty, but a great deal of work has been done to ensure completely safe operation of all parts. Safety ignition devices for oven and grill burners have been obligatory since January, 1960, and agreement has been reached between the German Gas Association and the appliance makers that, from 1st January, 1963, only completely safe domestic gas appliances shall be marketed. In addition, detachable safety tubing for connecting gas cookers was introduced a year ago. The Author gives detailed descriptions, with illustrations, of the various types of safety ignition device—the bimetallic (mainly for oven burners), the thermo-electric (for hotplate, oven and grill burners), and other newer ignition and safety devices (mainly for semi-automatic and fully automatic cookers). A fully automatic gas cooker is described in which safe ignition can still be ensured even if the electricity supply fails. A great future is also foreseen for fully automatic gas space heating.

Change to Gas-fired Radiant Heat Increases Lehr Output 30 per cent.

L. Dubeau and G. S. Mendell. *Industrial Gas*, 1961, October, 8-9.

Previous attempts, by the largest manufacturer of glass containers in Canada, to increase production efficiency and capacity of lehrs using electric heating, resulted in excessive breakage and rejects. By substituting a gas-fired, radiant-burner roof and lines of small perforated-refractory blue-flame gas burners for the electrical elements below the belt, the production capacity of the lehr has been increased 30 per cent, and it is estimated that this could be raised to 50 per cent, if required. Annealing rejects have been eliminated, and rigid quality standards are met without fail. Since more than one-half of the heat is now applied from above instead of entirely through the belt, repair and replacement costs on the latter have been greatly reduced. The radiant roof heaters burn a controlled gas/air mixture against a special cup-shaped refractory surface that beams radiant heat to the glass containers, whereby the latter are brought to annealing temperature faster, permitting an increase in belt speed and in the number of containers annealed per hour. The Authors describe in detail the annealing cycle of the converted lehr.

Exhibition of Paintings and Drawings

The Council of the Institution has approved a suggestion that an exhibition of paintings and drawings by members of the gas industry be held at the time of the Annual General Meeting in 1962, and a small Committee has been formed to look further into the matter.

The Exhibition would probably be held at the offices of the Institution at 17, Grosvenor Crescent, London, S.W.1, and, depending upon the response to this original proposal, would be repeated every second or third year.

It is very desirable, at this early stage, to obtain some idea of the likely number of works that would be available, and it is hoped that everyone who is interested in the subject will support this first venture.

Paintings and drawings in any medium would be welcomed and exhibited without competitive selection.

It would be of the greatest assistance to the organizers if those willing to exhibit would kindly write **now**, stating the number and sizes of works available, to the following :— Ralph Pollard, Esq., Lee Cottage, Weston Green, Thames Ditton, Surrey. (Telephone : Emberbrook 6329.)

District Sections and Affiliated Associations

(a) Notes of Meetings

DISTRICT SECTIONS

London and Southern Section

The first meeting of the 1961-62 session was held at the Caxton Hall, Westminster, on Tuesday, 17th October, 1961.

Before the commencement of the Annual General Meeting, Mr. J. A. Hepworth, B.Sc., the retiring Chairman, introduced Mr. R. M. Farror as the Chairman for the forthcoming session. Following tradition, Mr. Hepworth was presented with a memento of his year of office. In making the presentation, Mr. Farror complimented Mr. Hepworth on a successful session that had included some very interesting papers, and said he himself would do his utmost to maintain the high tradition of the office of Chairman.

The minutes of the 10th Annual General Meeting, held on 18th October, 1960, were read and, on the proposal of the Chairman, confirmed. The report of the management committee, and the statement of accounts, for the 1960-61 session, which had been circulated to all members, received no comments and were formally adopted. The Management Committee's nominees for new officers of the Committee were unanimously approved and in consequence Mr. G. G. Warne was elected Junior Vice-Chairman and Mr. C. H. Belben, Mr. L. J. Clark, B.E.M., and Mr. C. A. Deas were elected as Committee members in place of the retiring members.

On behalf of the Institution, the Chairman presented certificates of Associate Membership of the Institution to Mr. K. F. A. Carrier, Mr. C. Clayton, Mr. A. A. Dunford, Mr. A. M. Haken and Mr. D. J. Huggett.

Mr. W. E. Medhurst, Assistant Chief Distribution Engineer, South Eastern Gas Board, then presented a paper entitled "The Organization and Operation of a Distribution Department".

Before opening the discussion, the Chairman thanked Mr. Medhurst for preparing such an interesting and comprehensive paper on a subject on which the Section had not had a paper for some years. He mentioned that the large expenditure by the gas industry since vesting day on gas supply and storage indicated the importance of this aspect of gas engineering. Also, the wholesale closing down of a large number of works had increased not only the work, but, in addition, the responsibilities of distribution departments.

A vote of thanks to Mr. Medhurst was proposed by Mr. D. J. O. Bath.

South Western Section

The Annual General Meeting of the South Western Section was held at Bristol on Wednesday, 18th October, 1961, with some 65 members and guests present. The proceedings were presided over by Mr. T. W. Jackson, Chairman. Prior to the meeting, luncheon was taken at the Royal Hotel, Bristol.

Mr. Jackson welcomed Mr. A. G. Higgins, M.Sc. (Secretary, The Institution of Gas Engineers) and Mr. C. H. Chester, C.B.E. (Chairman, South Western Gas Board).

The following officers were elected for the ensuing year :—Chairman, Mr. R. J. Williams ; Senior Vice-Chairman, Mr. T. W. Clapham ; Junior Vice-Chairman, Mr. W. Dougall ; District Member of Council, Mr. T. W. Jackson ; Honorary Secretary, Mr. F. C. Gay, M.B.E. ; Honorary Treasurer, Mr. T. A. Nicholson.

The following were elected as ordinary members of the Management Committee :—Mr. F. Tweedle, Mr. R. Pratt and Mr. P. J. Whitehead.

It was unanimously agreed that the annual donation to the Institution Benevolent Fund should be increased to 35 guineas, and Mr. A. G. Higgins thanked the Section for its continued generosity. The luncheon collection for the Fund on this occasion totalled £6 7*s.* 10*d.*

Presenting the annual report and accounts, Mr. F. C. Gay, M.B.E., said the Section had had a very high standard of papers at its meetings, and these had been very well supported. The accounts showed a smaller adverse balance than in the previous year, and the increased subscription should put the finances straight in the next two or three years. The annual report and accounts for 1960-61 were unanimously adopted.

Mr. Philip Albert Edwin Wallis, Assoc.I.Gas E., was elected a member of the Section.

The Chairman presented to Mr. G. F. Williames a certificate of Associate Membership of the Institution.

The retiring Chairman, Mr. T. W. Jackson, presented his Address, which covered "highlights" in the Section's 10 years' history, and touched upon a few of the problems facing the gas industry today.

Mr. C. H. Chester, C.B.E., congratulated Mr. Jackson on his excellent Address, and expressed appreciation of the valuable work undertaken by him.

Mr. Jackson introduced Mr. R. J. Williams the new Chairman of the Section, who expressed appreciation of the honour conferred upon him and congratulated the retiring Chairman on having achieved a very high standard in his programme during the past session.

The incoming Chairman then presented Mr. Jackson with his Past-Chairman's Badge to commemorate his year of office, together with a framed 16th Century map of Gloucestershire.

Manchester and District Section

Some 50 members joined the Chairman, Mr. N. C. Sturrock, on a visit to Car House Works, Rotherham, at the invitation of the East Midlands Gas Board on 29th September, 1961. The members were entertained to luncheon, when they were welcomed by Mr. E. H. Harman, O.B.E., B.Sc., Deputy Chairman of the Board, in the unavoidable absence of the Chairman, Mr. R. S. Johnson, M.B.E., T.D., M.A., LL.B., who was unable to be present. The attendance had been

adversely affected at the last moment by the untimely death of Mr. T. W. Pickthall, and Mr. Harman said how sorry he was to learn of the loss the Section had just sustained.

He referred to the developments at the gas works, with which he had been closely connected, and to the success that was attending the Board's efforts to increase sales of gas for space heating. Contrary to expectations, instead of this business merely producing a peak load in winter, it had apparently had the effect of lifting the load curve throughout the year.

Mr. Harman referred to the fact that, although most of the members of the Institution employed by the East Midlands Gas Board were members of the Midland Section, quite a number still retained an interest in the Manchester and District Section, and the Board was glad to have this opportunity to welcome the Section into its Area.

Mr. Sturrock thanked Mr. Harman and, through him, the Chairman of the Board, for the invitation that had been extended to the Section.

During the afternoon, members inspected the works and were impressed by its cleanliness and the orderliness of the activities of the plant—matters that were referred to in the vote of thanks made by the Chairman of the Section when the party was subsequently entertained to afternoon tea. Mr. C. C. Wood, General Manager of the Sheffield and Rotherham Division, who replied, said how much pleasure it had given to him and his staff to welcome the Section on this occasion.

JUNIOR GAS ASSOCIATIONS

London and Southern Junior Gas Association

(*a*) The first meeting of the 1961-62 session, held at the Westminster Technical College on Friday, 6th October, 1961, opened with Mr. P. J. Savage, the retiring President, in the chair. Mr. Savage asked members to stand as a token of respect to Mr. T. W. Pickthall, Chairman of the British Junior Gas Association Joint Council, who had died as a result of an accident.

Mr. Savage then presented the President's Prize for 1960-61 to Mr. R. J. Langford for his paper on "The Planning of Small Holder Stations"; and the Milbourne Prize to Mr. P. R. Chance and Mr. E. W. G. Dance for their paper on "Laboratory and Field Experience in Warm-air Heating".

He then congratulated Mr. M. J. Birch and Mr. H. T. Lockwood on their award of the Society of British Gas Industries' Silver Medal for their paper on "Catalytic Gasification at Portsmouth and Southampton".

Mr. Savage then introduced the new President, Mr. C. H. Townsend (Wandsworth), who delivered his Presidential Address.

Mr. E. A. K. Patrick, B.Sc., Senior Vice-President, proposed a vote of thanks for the Address.

(*b*) On the afternoon of 12th October, 1961, a party of more than 50 members of the Association were guests of the South Eastern Gas Board on the occasion of a visit to the East Greenwich Works of the Board. The works occupies more than 90 acres alongside the Thames at Blackwall Point. Having been considerably developed over the past 10 years, the works was used by Mr. C. H. Townsend in his Presidential Address to the Association as an illustration of development during this period.

The maximum capacity (107 mill. ft³/d) of the works is made up of gas from four batteries of coke ovens, six benches of horizontal retorts, 10 carburetted water gas sets equipped for reforming light feedstock, and the butane/air installation.

In parties of not more than eight, the visitors were taken over the longer distances in personnel carriers, and the routing and timing of the whole visit was planned by the Board's Work Study Department. In this way, a wide picture was obtained of the works as a whole. Particular interest was shown in the instrumentation and automatic controls of the butane plant, and also in the control-room.

After tea, taken in the Works Institute, Mr. B. C. Hinwood, Senior Station Engineer to the Board, in welcoming the visitors paid tribute to the work of the brothers Livesey and their far-sighted planning so many years ago.

Mr. C. H. Townsend, thanking Mr. Hinwood and Mr. A. J. L. Gamper, Station Engineer at East Greenwich, for hospitality, and the guides and staff concerned for their efforts, said that it was fitting that in this, its Diamond Jubilee session, the Association should again visit Greenwich as it had done in both the Silver and Golden Jubilee sessions. Before dispersing, the party continued discussion around the interesting scale model of the works, which was on display.

(*c*) The Association held its annual reception and dance at Caxton Hall, Westminster, S.W.1, on 3rd November, 1961, when 170 members and guests were received by Mr. C. H. Townsend (President) and Mrs. Townsend.

In the course of the evening, Mr. Townsend formally welcomed the official guests: Mr. R. N. B. D. Bruce, O.B.E., T.D., B.A., B.Sc., Chairman, South Eastern Gas Board; Mr. C. Stott, M.C., Chief Engineer, South Eastern Gas Board; Dr. A. E. Haffner, B.Sc., member and Chief Engineer, Southern Gas Board, and Senior Vice-President, The Institution of Gas Engineers; Mr. A. G. Higgins, M.Sc., Secretary, The Institution of Gas Engineers; Mr. Ben. Morgan; Mr. R. M. Farror and Mr. J. T. Veryard, B.Sc., Chairman and Honorary Secretary respectively of the London and Southern Section of the Institution, and Mr. A. H. Savill, Senior Production Engineer, South Eastern Gas Board, and Past-President of the Association.

Replying for the guests, Mr. Bruce said that 3rd November, 1961, would be remembered in the history of the British gas industry, for the Government had that morning approved and authorized The Gas Council's application to import Saharan methane as a feedstock for the manufacture of town gas.

Arrangements for the dance were in the hands of Mr. E. G. Duggin and Mr. T. J. Batsford, the Association's Honorary Assistant Treasurer.

Manchester District Junior Gas Association

The Association's 63rd Annual General Meeting and President's Day was held at Manchester on 21st October, 1961.

Prior to the meeting, Mr. J. A. Taylor, President of the Association, and about 90 members visited the Industrial Equipment Exhibition at the City Hall, where particular attention was paid to the comprehensive exhibit arranged by the North Western Gas Board.

At the Annual General Meeting, which was held at the North Western Gas Board's Industrial Development Centre, Mr. G. Gilden, Production Engineer of the Board's Manchester Group, welcomed members on behalf of the Board and apologized for the absence of Mr. D. P. Welman and Mr. W. Hodkinson, O.B.E., Chairman and Deputy Chairman respectively of the North Western Gas Board, who were both indisposed. Mr. Gilden referred to the pleasure given to the Manchester Group by the selection of three Manchester men to fill the office of President of the Association in consecutive years—Mr. J. A. Taylor (retiring President), Mr. W. F. Howell (President-Elect) and Mr. D. W. Kay (Senior Vice-President).

During the meeting, Mr. J. A. Taylor presented the Col. W. M. Carr Memorial Prize to Mr. N. Hoyle, and the F. Johnston Prize, for the best short paper, to Mr. J. D. Wilson.

Mr. John West, on behalf of the Society of British Gas Industries, presented the Society's Silver Medal for 1960 to Mr. W. Moorcroft.

Mr. Taylor then installed Mr. W. F. Howell as President for the session 1961-62, after which the new President presented Mr. Taylor with a Past-President's Medallion.

Mr. Howell delivered his Presidential Address, for which a vote of thanks was given on the proposition of Mr. D. W. Kay (Senior Vice-President), seconded by Mr. G. Brook (Junior Vice-President).

Following the meeting, members were entertained to tea at the invitation of Mr. A. H. Nicholson, General Manager of the Manchester Group.

Mr. J. R. Teale expressed the thanks of the Association to Mr. Nicholson and all those who had been responsible for organizing an extremely successful day, to which Mr. Gilden briefly responded.

Midland Junior Gas Association

The President, Mr. R. W. Nelson, was in the Chair and 75 members were present at a meeting held in the Staff Mess Room of the West Midlands Gas Board, at Birmingham, on Tuesday evening, 10th October, 1961.

The President announced that the Milbourne Prize for the session 1960-61 had been awarded to Mr. C. Halstead for his paper on " The Economic Use of Steam on Two Large Gas Works ".

The Honorary Secretary (Mr. W. E. Dobson) announced that the proposed paper by Mr. R. Hutchinson would be presented on 14th November, not the 21st as indicated in the programme.

A strong appeal was made by the President for members to consider preparing suitable papers for consideration by The Institution of Gas Engineers for presentation at the Institution's 1963 Annual General Meeting.

The President said that it gave him pleasure to announce that Mr. C. H. Leach, M.A., Chairman, West Midlands Gas Board, had accepted Honorary Membership of the Association.

Mr. Nelson said that before installing the new President into office he would like to thank members of the Council and of the Association for the support they had given to him during his year of office. The President then introduced Mr. A. V. Wainwright, Engineer and Manager, Shrewsbury District, West Midlands Gas Board, and formally installed him in the Chair and presented him with the President's Badge of Office. The newly elected President then submitted his Presidential Address, which was well received. Mr. B. G. H. J. Hawkings, Junior Vice-President, proposed a hearty vote of thanks to the President for his Address, which he thought all members had found most interesting.

At the invitation of the President, light refreshments were served at the conclusion of the meeting.

(b) Abstracts and Discussions of Papers

Copies of the complete papers abstracted below are available for consultation or loan in the Library of the Institution.

DISTRICT SECTIONS

London and Southern Section

"The Organization and Operation of a Distribution Department",

by W. E. Medhurst, M.I.Gas E., Assistant Distributing Engineer, South Eastern Gas Board.

Paper delivered in London on 17th October, 1961.

In May, 1956, Mr. R. Prince and Lt.-Col. J. A. Gould, M.C., presented to The Institution of Gas Engineers a paper (*Communication* 485) describing the organization set up by the South Eastern Gas Board to rationalize the production and transmission of gas. At that date, gas was being manufactured at 44 works and there were 84 holder stations. Since then, integration has continued to take place, with a corresponding reduction in the number of manufacturing stations, and there are at the present date 14 works and 105 holder stations.

In 1957, the control of holder stations and the transfer of gas was made the responsibility of the Board's Distribution Department, and considerable thought was given to the development of a divisional organization that could carry out the additional duties in an efficient manner.

It was decided to divide the department into three main functional lines, each controlled by the following specialist officers :— (*a*) administrative assistant, (*b*) mains and services engineer, (*c*) supply engineer.

These officers, together with the Assistant Distributing Engineer, form a compact team working under the control of the Distributing Engineer.

Each officer must be well in the picture regarding the general operation and policy of the department, and the administrative assistant must see that this object is achieved.

In order that a common policy shall be operated throughout the Board's Area, meetings are held under the chairmanship of the Chief Distribution Engineer, where the senior officers of the

department discuss items of common interest and receive advice on general policy.

Each administrative assistant has a small staff at Divisional Headquarters, who deal with expenditure control, statistics, correspondence and general paper work concerned with distribution matters.

The mains and services engineer is responsible for all main and service laying, which may be carried out by direct or contract labour in his Division. He has on his staff distribution supervisors who are responsible for work in particular sections of the Division. Each supervisor has a clerk who carries out various duties such as issuing job cards, keeping records, preparing time-sheets and arranging transport, *etc.*

A radio transmitter/receiver in the office ensures close contact with the supervisor and foremen as well as with the various vehicles fitted with similar equipment.

It is important that the mains and services engineer should keep in close contact with similar engineers of other undertakings, as considerable savings can be made by careful planning of work in conjunction with these people.

The supply engineer is responsible for all matters relating to the storage, transmission and distribution of gas in his Division. He has on his staff gasholder stations foremen who are craftsmen and responsible for groups of holder stations. Maintenance is carried out using a small maintenance gang attached to each foreman's area, or by employing the services of the staff at a major manufacturing works, or by employing contractors.

Gas Control

More than 70 per cent of the gas output in the South Eastern Gas Board's five London Divisions is pumped through transmission, mains, and the balance between gas made at each works and gas pumped must be controlled in order to achieve economic operation.

An Area Control at the Board's Headquarters plans the divisional programmes of gas to be made, gas to be received or transferred, and the total gas stock levels to be held. These figures are agreed with the Sub-Area Control in each Division, where the internal divisional transmission rates and the stocks to be held at individual holder stations are programmed in conjunction with the supply engineer.

Mobile Booster and Polythene Rider

Since more than 70 per cent of the Board's gas output is pumped through transmission and feeder mains, it follows that a failure of one of these could be very serious.

A rider across the affected section is only a part answer unless the cross-sectional area of the rider is equal to that of the main, or, alternatively, some means of restoring the pressure drop across the rider is provided. In order to meet this latter need, a diesel-driven booster has been mounted on a trailer suitable for towing. A half-mile of 62in. i.d. polythene tubing in 16-ft lengths is available for use with the mobile booster. The tubing is stacked in steel crates for lorry mounting, and the whole equipment can be made ready to proceed to the site of emergency in a very short time.

Special manifolds on the booster enable up to six tubes to be laid in parallel.

Modification to Gasholder Stations

When considering a reconstruction programme for a holder station, three main points have been kept in mind. They are (*a*) safety of personnel and plant, (*b*) saving in maintenance and operating costs, and (*c*) security of operation.

Maintenance costs must be considered when designing the layout; it is too late to try to reduce these costs after the reorganization is completed.

The erection of plant in the open will almost completely eliminate the chance of an explosion due to leaking gas. Governor pits have been designed with a maximum depth of 4 ft 6 in., so that a man standing on the floor of the pit will have his head and shoulders above ground.

Good maintenance must play an important part in the security of operation of a holder station. Safety governors should be installed between the inlet and outlet mains on the station, and should be set to operate at a pressure that follows just below that worked on the station governor.

Modification of Old Kent Road and Kennington Holder Stations

Both these stations together have a maximum day of approx. 43 mill. ft^3 of gas, Old Kent Road receiving gas from the Board's major works at East Greenwich. Pumping plant at Old Kent Road can transmit 24 mill. ft^3 of gas to Kennington and Rotherhithe holder stations.

Kennington holder station is also connected to the North Thames Gas Board's transmission and low-pressure distribution system, and the connexions are arranged so as to enable gas to be supplied from either Board to the other.

Diesel- and electrically-driven transmission boosters and district boosters have been erected at Old Kent Road, the boosters being in the open air with the prime movers situated in a simple framed building.

The boosters at Kennington are situated in the open, with the diesel prime mover covered with a sectional aluminium cubicle lined with a sound-absorbing material.

All pits for governors and non-return valves are covered with aluminium covers that can be lifted or drawn off by one man.

In view of the importance of these two stations, they are manned throughout the 24 h, instructions regarding stock, gas to be received and pumped, being given by the Sub-Area Control at East Greenwich.

Discussion

Mr. A. D. L. Copp (Distributing Engineer, North Thames Gas Board) complimented the Author on the number of points discussed, and the admirable diagrams. He agreed that works that had been shut down were not readily adaptable for distribution stations; more often than not, no plant other than the holders was of much use. On the need for specialist engineers, the North Thames Gas Board relied on the specialists on the production side, and on the organization of staff his own Board's system differed somewhat with rather fewer numbers. He referred to the necessity of appraising all pumping costs, which, although often individually small, could add up to quite a considerable total. He considered the use of polythene tubes most ingenious, and asked whether they were likely to collapse under reduced pressure conditions. He finally referred to the collaboration between the North Thames and the South Eastern Gas Boards whereby inter-communicating connexions terminated in valves that could be operated by either Board.

Mr. B. Wilkinson (Distribution Engineer, Eastern Gas Board) agreed that it was essential to separate the production of gas from its distribution, but said that in the Eastern Gas Board some locations were so remote from the headquarters that it was necessary for some of the distribution staff to be under the control of the local production engineer. He raised a number of questions regarding the keeping of records, and, in regard to the siting of compressors, felt that some form of cheap building was desirable for it led to better maintenance.

Mr. G. G. Warne (Deputy Distribution Engineer, Southern Gas Board) drew attention to the change in the organization described in the paper in comparison with that given to the Institution by Mr. Prince and Lt.-Col. Gould, and maintained that, whatever the form of administration might be, there must be the same common purpose of providing a thoroughly efficient consumer service. He made a number of comments on the function of the clerical section. On the operational side, he wondered if jet boosters had any real value since there appeared to be the necessity for alternative standby plant and drew attention to the necessity for adequate maintenance and testing of safety governors. He finally questioned whether electrically-powered booster plant is cheaper than diesel-driven plant.

Mr. D. J. O. Bath said the Author was to be congratulated on the lucid description of the organization and operation of his Board's Distribution Department. He considered the rearrangement of the plant at Old Kent Road and Kennington Works to be splendid examples of concentrated design offering ease of operation and maintenance, and considered the use of plastic tubes for emergency gas connexions to be of particular interest.

Mr. Medhurst replied to the many questions raised during the discussion.

JUNIOR GAS ASSOCIATION

London and Southern Junior Gas Association

Presidential Address, *by C. H. Townsend, B.Sc., A.C.G.I., M.I.Mech.E., M.I.Gas E., A.M.Inst.F., Station Engineer, Wandsworth Works, South Eastern Gas Board.*

Address delivered in London on 6th October, 1961.

In view of the fact that it is the Diamond Jubilee Year of the Association, Mr. Townsend's Address reviews the progress of the gas industry in the last 10 years, carrying on from the review of the first 50 years made by Mr. J. D. C. Woodall in the Golden Jubilee Year.

Developments in gas manufacture are examined, starting with the effects of rationalization following nationalization, leading to the development of a small number of large works. The effects of the considerable price increases of coal, and the changes in the coke market conditions as they affect carbonization, are discussed. The development of more reactive cokes for domestic use, such as " Cleanglow ", " Gloco " and " Sebrite ", are mentioned.

The greater use of carburetted water gas plants and the effect on costs of the increasing price of coke are looked at, and emphasis is placed on the considerable assistance to the process, both in output and costs, due to the development of the use of light petroleum distillates. The measures necessary to ensure naphthalene removal with the modified process are mentioned.

The search for alternative processes and primary fuels, and the progress made in this direction are examined.

Developments at the East Greenwich Works of the South Eastern Gas Board are quoted as an example of the development of large works, and a description of the plant is given with some details of operating experience, particularly in relation to steam and power utilization.

The point is made that the East Greenwich Works has been developed from a capacity of 36 mill. ft³/d in 1950 to 107 mill. ft³/d in 1961, retaining only 18 mill. ft³/d capacity of the original plant. The effect of these developments on costs of manufacture are set out, showing that part of the increased cost of coal has been contained by higher efficiencies and that works operating costs have been reduced in spite of the rise of the index for salaries and wages from 100 to 177 in the period reviewed.

Mention is made of the considerable impact on distribution technique of integration schemes and the higher and higher pressures used to distribute gas, together with greater use of instruments and automatic controls.

In utilization, the development of new appliances, especially for heating, are reviewed, and the importance on the effective cost of gas used in appliances of higher efficiency is underlined.

Turning to the effect of change on people, the increase of specialization is referred to, and the President comments on the benefits of residential management courses and the forum of the " Juniors " as places for specialists to meet and get to know each other and each other's problems.

From this study of reaction to change, the conclusion is drawn that the industry's chances of survival are good if the people in it have faith and persevere as previous generations have done.

In this Issue...

J. B. Lane

J. B. Lane is the third generation in the tar distilling family business of Lane Brothers (Tar Investments), Limited. His father, J. B. Lane, senior, was a prominent member of the industry until his death in 1947. His uncle, J. R. Lane, was the other half of the Lane Brothers, and was a well known member of the tar industry until his death in 1943. The Lane Brothers business is still a fact, the present other half being M. J. R. Lane, a younger brother.

Educated at Rossall School, 1938-43, with a brief spell at Clare College, Cambridge, before joining the Army in 1943, Mr. Lane served in the Royal Engineers, becoming a Lieutenant, R.E. He served for three years in the Far East, and was demobilized in the Spring of 1947. He resumed at Clare College, Cambridge, in the Autumn Term, but on the death of his father, in December, cut short his academic career and joined the family business.

The family business trades as three main companies, the Lincolnshire Chemical Company, Limited, The Normanby Park Tar Supply Company, Limited, and the North Western Co-operative Tar Distillers, Limited, the last named being part of Lancashire Tar Distillers, Limited. From 1947 to 1953, Mr. Lane remained with the Lincolnshire Chemical Company and Normanby Park Tar Supply Company, Limited, and became managing director in 1951. In 1954, he transferred to Lancashire Tar Distillers as technical director, and became joint managing director in 1956. His brother took over as managing director of the other companies in 1954. In 1957, on the death of Col. W. M. Carr, the then Chairman, he became chairman of Normanby Park Tar Supply Company, Limited ; Lincolnshire Chemical Company, Limited, and North Western Co-operative Tar Distillers, Limited, and in 1960, became vice-chairman of the newly formed Port Talbot Chemical Company, Limited. In addition, he is a director of the following firms :— Manchester Creosote and Storage Company, Limited; Lane Brothers (Tar Distillers), Limited ; R. Graesser, Limited.

Married, with two children, Mr. Lane lives at Bowdon, Cheshire. Though an ex-Rugby player, who played in 1949 for the North Midlands, and an ex-village cricketer; he now confines his recreational outdoor activities to an occasional bit of shooting.

Forthcoming Meetings

The following meetings have been arranged for the six weeks commencing 1st December, 1961:

(*a*) District Sections

THE INSTITUTION OF GAS ENGINEERS—EASTERN SECTION

6th December — Joint Meeting with the Eastern Junior Gas Association. Paper, "The Background of Industrial Relations in the Gas Industry", by A. B. Badger, M.A., Ph.D. (The Institution of Gas Engineers, 17, Grosvenor Crescent, London, S.W.1.)

THE INSTITUTION OF GAS ENGINEERS—NORTH OF ENGLAND SECTION

7th December — Joint Meeting with the Northern Junior Gas Association. Showing of The Scottish Gas Board's film of the crossing of the River Forth, with introduction and commentary by R. C. Emmony. (2.30 p.m. The Lecture Room, News Theatre, Newcastle-upon-Tyne.)

THE INSTITUTION OF GAS ENGINEERS—SOUTH WESTERN SECTION

6th December — Paper, "A Technical Exercise in Functional Inter-dependence", by Dr. M. J. F. Olden and P. G. Jackson. (Bristol). *Not Gloucester as stated in the November issue of the Journal.*

(*b*) Junior Gas Associations

EASTERN

6th December — Joint Meeting with the Eastern Section, The Institution of Gas Engineers. Paper, "The Background of Industrial Relations in the Gas Industry", by A. B. Badger, M.A., Ph.D. (The Institution of Gas Engineers, 17, Grosvenor Crescent, London, S.W.1.)

LONDON AND SOUTHERN

1st December — Paper, "The Safe Operation of Mechanical Handling Plant", by R. W. Cutts. (6.30 p.m. Westminster Technical College, Vincent Square, London, S.W.1.)

14th December — Visit to J. Lyons and Company, Limited, Cadby Hall, London, W.14.

15th December — Paper, "Retrospect on American Visit", by S. B. Scott. (6.30 p.m. Westminster Technical College, Vincent Square, London, S.W.1.)

5th January — Short Paper Meeting. "Transmission of Propane/Butane or similar Liquids by Pipeline", by J. B. Kempton. "An Inferential Meter", by C. B. Henshilwood. "Gas Technik Purifiers", by W. J. Hayward. (6.30 p.m. Westminster Technical College, Vincent Square, London, S.W.1.)

MANCHESTER

9th December — Short Papers for F. Johnston Prize. (Manchester Industrial Development Centre.)

MIDLAND

12th December — Short-paper Evening. (West Midlands Gas Board, Staff Mess Room, Council House, Birmingham.)

9th January — Paper, "Recent Developments in Retort House Practice at Saltley", by H. B. Coan. (West Midlands Gas Board, Staff Mess Room, Council House, Birmingham.)

NORTHERN

7th December — Joint Meeting with the North of England Section, The Institution of Gas Engineers. (Newcastle.)

19th December — Short-paper meeting for the President's Prize. (2.30 p.m. Gas Showrooms, Blyth.)

SCOTTISH (WESTERN)

9th December — Joint Meeting with the East of Scotland Junior Gas Association. (3 p.m. Albany Street Social Club, Edinburgh.)

WALES AND MONMOUTHSHIRE

5th December — Visit to Grangetown Gas Works of the Wales Gas Board. Paper, "Recent Developments at Cardiff", by J. B. Currier. (Cardiff.)

WESTERN

13th December — Paper, "Central Heating by Gas—Design Considerations", by E. I. Luscombe. (Plymouth.)

(*c*) Kindred Bodies

Members of The Institution of Gas Engineers who wish to attend any of these meetings, but are not members of the bodies concerned, should apply to the appropriate Secretary at the address indicated.

THE CHEMICAL SOCIETY
(Burlington House, London, W.1.)

5th December	Paper, " Some Recent Observations on the Activity of Metal Catalysts ", by Prof. C. Kemball. (4.30 p.m. The University, Leicester.)
8th December	Joint Meeting with the Royal Institute of Chemistry, the Society of Chemical Industry and the Society for Analytical Chemistry. Paper, " The Structure of Natural Products by Direct X-Ray Analysis ", by Prof. J. Monteath Robertson. (7.15 p.m. The Royal College of Science and Technology, Glasgow.)
14th December	Liversidge Lecture. " Stereospecific Polymerization ", by Prof. C. E. H. Bawn, C.B.E. (7.30 p.m. The Lecture Theatre, The Royal Institution, Albemarle Street, London, W.1.)
10th January	Paper, " Precision Balances : History, Development and Techniques ", by A. O. Brooks, with a film " The Balance and its Use ". (Chemistry Department, University, Aberdeen.)

THE INSTITUTE OF FUEL
(18, Devonshire Street, Portland Place, London, W.1.)

5th December	Paper, " Recent Developments in the Heating of Domestic Premises ", by P. T. Loader. (6.30 p.m. The Gas Showrooms' Lecture Theatre, Nottingham.)
7th December	Paper, " Developments at Garston Works ", by W. S. Hubbard. (7 p.m. Liverpool Engineering Society's Rooms, 9, The Temple, 24, Dale Street, Liverpool, 3.)
7th December	Paper, " The Benson Boiler ", by W. P. C. Ungoed. (6.30 p.m. Birmingham Exchange and Engineering Centre, Stephenson Place, Birmingham, 2.)
13th December	Joint Meeting with the National Society for Clean Air. Paper, " Shell Boiler Developments ", by G. Gunn. (6.30 p.m. The Engineers' Club, Albert Square, Manchester, 2.)

THE INSTITUTE OF PHYSICS
(47, Belgrave Square, London, S.W.1.)

6th December	Paper, " The Science Masters' Association Syllabus ", by E. W. Tapper. (5.30 p.m. 47, Belgrave Square, London, S.W.1.)
8th December	Paper, " The Correlation of Elastic Constants in Single Crystals and Polycrystals ", by M. F. Markham. (5.30 p.m. 47, Belgrave Square, London, S.W.1.)
11th December	Paper, " Static Electricity ", by P. S. H. Henry. (5.30 p.m. 47, Belgrave Square, London, S.W.1.)
12th December	Meeting on " Strain Gauges ". (2.30 p.m. 47, Belgrave Square, London, S.W.1.)
12th December	Paper, " Integrated Electronics ", by J. W. Granville. (5.30 p.m. 47, Belgrave Square, London, S.W.1.)
13th December	Meeting on " Ferromagnetism and Antiferromagnetism ". (11 a.m. 47, Belgrave Square, London, S.W.1.)
5th January	Symposium on " Some Aspects of Vacuum Science and Technology ". (Imperial College of Science and Technology, South Kensington, London, S.W.7.)

THE INSTITUTION OF ELECTRICAL ENGINEERS
(Savoy Place, London, W.C.2.)

12th December	Paper, " Diversity Reception and Automatic Phase Reception ", by L. Lewin. (5.30 p.m. Savoy Place, London, W.C.2.)
19th December	Paper, " The Use of Controllable Semi-conductor Rectifiers in Machine Control ", by K. G. King. (5.30 p.m. Savoy Place, London, W.C.2.)

THE INSTITUTION OF ENGINEERS-IN-CHARGE
(Honorary Secretary : A. T. Penn, Esq., 13, Grange Close, Edgware, Middlesex.)

13th December	Paper, " The Hot Dip Galvanizing Process ", by M.H. Davies. (6.30 p.m. Caxton Hall, London, S.W.1.)
10th January	Paper, " Free Piston Engines and their Application ", by L. Copeland Watts. (6.30 p.m. Caxton Hall, London, S.W.1.)

THE INSTITUTION OF HEATING AND VENTILATING ENGINEERS
(49, Cadogan Square, London, S.W.1.)

5th December	Paper, " Experience in America of Heating and Air Conditioning Practice ", by C. S. K. Benham, O.B.E. (6.30 p.m. The Institute of Mining and Mechanical Engineers, Neville Street, Newcastle-upon-Tyne.)
6th December	Paper, " Some Aspects of the Work of the Heating and Ventilating Research Association ", by N. S. Billington. (6.30 p.m. The College of Art and Crafts, Waverley Street, Nottingham.)
8th December	Paper, " Some Important Aspects of Oil-firing as Applied to Sectional and Shell Boilers ", by N. Charlesworth. (6.30 p.m. The Engineers' Club, Albert Square, Manchester, 2.)
12th December	Paper, " Fans : Types and Uses ", by J. W. Stammers. (6.30 p.m. The Engineering Centre, Stephenson Place, Birmingham.)
14th December	Paper, " Fire Protection in Buildings ", by E. A. Hibbitt, M.B.E. (6 p.m. The Institution of Mechanical Engineers, 1, Birdcage Walk, London, S.W.1.)

THE INSTITUTION OF MECHANICAL ENGINEERS
(1, Birdcage Walk, London, S.W.1.)

6th December	Paper, "Lubrication Oil Filter Testing", by K. E. Buckman and S. W. Kemp. Paper, "A Standard Contaminant for Oil Filter Testing", by A. T. Weare. (6 p.m. 1, Birdcage Walk, London, S.W.1.)
8th to 12th January	International Heat Transfer Conference. Repeat of Conference held in Boulder, Colorado, U.S.A., from 28th August to 1st September, 1961. (Lecture Hall, Central Hall, Westminster, London, S.W.1.)

THE INSTITUTION OF PLANT ENGINEERS
(2, Grosvenor Gardens, London, S.W.1.)

5th December	Paper, "The Fuel Cell", by F. T. Bacon. (7 p.m. Royal Society of Arts, John Adam Street, Adelphi, Strand, London, W.C.2.)
5th December	Joint Meeting with the Institution of Production Engineers. Paper, "Non-destructive Testing", by F. Hinsley. (7.30 p.m. Conference Room, F. Perkins, Limited, Peterscourt, Peterborough.)
7th December	Paper, "Electronic Control in Industry, with Particular Reference to Automation". (7.30 p.m. The Hotel Leofric, Coventry.)
7th December	Paper, "Repairs by the Metalock Process", by N. Tinwell. (7 p.m. Sherwood Room, County Hotel, Theatre Square, Nottingham.)
13th December	Paper, "Safety in Factories", by J. B. Taylor. (7.15 p.m. Grand Hotel, Broad Street, Bristol.)
20th December	Paper, "Molybdenum Disulphide as a Lubricant", by D. E. Burton. (7 p.m. South Eastern Gas Board Lecture Hall, 95, High Street, Rochester, Kent.)
21st December	Paper, "The Merits of Various Heat-generation and -distribution Systems," by R. A. Wiggans. (7.30 p.m. The Castle Hotel, Blackburn.)

THE INSTITUTION OF STRUCTURAL ENGINEERS
(11, Upper Belgrave Street, London, S.W.1.)

1st December	Paper, "Instability of Thin-walled Members", by P. S. Bulson. (6.30 p.m. Main Lecture Theatre, Lanchester Building, The University, Southampton.)
5th December	Paper, "Crane Erection Problems", by W. H. Arch. (7 p.m. The Institution of Engineers and Shipbuilders, 39, Elmbank Crescent, Glasgow.)
6th December	Paper, "The Plastic Design of Steel-framed Structures", by Prof. M. R. Horne. (6.30 p.m. The College of Science and Technology, Manchester.)
6th December	Paper, "Some New Developments and Applications of Prestressed Concrete", by H. Kaylor. (6.30 p.m. The Neville Hall, Newcastle-upon-Tyne.)
14th December	Paper, "The Chemical and Physical Effects of Aggressive Substances on Concrete", by P. E. Halstead. (6 p.m. Meeting Hall, 11, Upper Belgrave Street, London, S.W.1.)

THE PIPELINE INDUSTRIES GUILD
(7, Iddesleigh House, Caxton Street, London, S.W.1.)

12th December	Paper, "Measurement of Flow in Pipelines", by Dr. F. A. L. Winternitz. (6.30 p.m. Tudor Room, Caxton Hall, Caxton Street, London, S.W.1.)
9th January	Paper, "The Contractors' Point of View", by I. Bowler, O.B.E. (6.30 p.m. Tudor Room, Caxton Hall, Caxton Street, London, S.W.1.)

THE ROYAL INSTITUTE OF CHEMISTRY
(Honorary Secretary, London Section: G. C. Ackroyd, Esq., B.Sc. Fire Offices' Committee, 107, Cheapside, London, E.C.2.)

7th December	Paper, "Ultra-violet Spectroscopy", by T. M. Dunn. (6.30 p.m. Northampton College of Advanced Technology, St. John's Street, London, E.C.1.)

SOCIETY OF CHEMICAL INDUSTRY
(14, Belgrave Square, London, S.W.1.)

8th December	Paper, "The Structure of Natural Products by Direct X-ray Analysis", by Prof. J. Monteath Robertson. (7.15 p.m. The Royal College of Science and Technology, Glasgow.)
12th December	Paper, "Production of Low-sulphur Aromatics", by G. Claxton. (6 p.m. 14, Belgrave Square, London, S.W.1.)
14th December	Special Meeting on New Test Methods. "The Use of a Density Column in a Storage-stability Test for Tar/Bitumen Mixtures", by Dr. E. D. Tingle, N. Wright and M. Stewart. "The Examination of Preserved Timber by Gas—Liquid Chromatography", by Dr. D. McNeil. "The Immersion Wheel-tracking Test", by D. H. Matthews. "Pipeline Enamels: Laboratory and Field Methods of Test", by G. McHardy. "A Test for Ease of Spreading of Bituminous Mixtures for Roads", by A. Please and R. Hardman. (2.30 p.m. 14, Belgrave Square, London, S.W.1.)

THE SOCIETY OF INSTRUMENT TECHNOLOGY
(20, Queen Anne Street, London, W.1.)

4th December	Paper, "Problems of Flow Measurement", by H. E. Farrer. (7 p.m. Department of Fuel Technology, The University, Mappin Street, Sheffield.)
7th December	Paper, "Electronic Controllers", by A. H. Isaac. (7.30 p.m. The Cleveland Scientific and Technical Institution, Middlesbrough.)
12th December	Paper, "Design Application and Selection of Automatic Control Valves", by P. Stone. (7 p.m. Stanley Palace, Watergate Street, Chester.)

Library Accessions

The following are a few additions to the Library since the issue of the *Journal*, Vol. 1, No. 11, November, 1961:

12th Annual Reports and Accounts, 1960-61, of the Area Gas Boards.

12th Annual Report and Accounts, 1960-61, of The Gas Council.

British Ceramic Research Association :

Accurate Analysis of Certain Ceramic Materials—a New Method, by H. Bennett, and others. *Research Paper* No. 494.

Pink Centres and Black Cores. A Study of Firing and Oxidation Rates Relating to the Elimination of Pink Centres and Black Cores in Dust-pressed Tiles, by W. H. Holmes, and others. *Technical Note* No. 12.

Preparation, Properties and Structure of Silicon Nitride, by P. Popper and S. N. Ruddlesden. *Research Paper* No. 495.

British Standard 779 : 1961 (Specification for Cast Iron Boilers for Central Heating and Hot Water Supply).

British Standard 855 : 1961 (Specification for Welded Steel Boilers for Central Heating and Hot Water Supply).

British Standard 3401 : 1961 (Specification for Beryllium Copper Crinkle Washers).

British Standard 3405 : 1961 (Simplified Methods for Measurement of Grit and Dust Emission from Chimneys).

British Standard 3406: Part 1: 1961 (Methods for the Determination of Particle Size of Powders, Part 1 : Sub-division of Gross Sample Down to 0·2 ml).

British Standard 3409 : 1961 (Specification for Tapping of Holes to Receive Wire Thread Inserts (UNC and UNF Threads).

British Standard 3410 : 1961 (Specification for Metal Washers for General Engineering Purposes).

Combustion, Flames and Explosions of Gases, by B. Lewis and G. von Elbe. 2nd ed. (Academic Press, 1961.)

Determination of Total Sulphur by the Rhône-Poulenc Method, by J. Pentelow. *Benzole Producers, Limited, Research Paper* 4-61.

Gas and Air Compression Machinery, by L. F. Scheel. (McGraw-Hill, 1961.)

Ground Storage of Liquefied Gases. (Conch Methane Services, Limited, 1961.)

Industrial Water Treatment Practice, edited by P. Hamer and others. (Butterworths, 1961.)

Long-term Demand for Scientific Manpower. (Advisory Council on Scientific Policy, Committee on Scientific Manpower, Statistics Committee, 1961.)

Ministry of Power Statistical Digest, 1960.

7th Progress Survey with Report and Accounts of the National Industrial Fuel Efficiency Service for the period 1st April, 1960, to 31st March, 1961.

Report from the Select Committee on Nationalized Industries : The Gas Industry, Vol. 1 : Report and Proceedings ; Vol. 2 : Minutes of Evidence, Appendices and Index. 1961.

Soil Adsorption of Odorant Compounds, by P. B. Tarman and H. R. Linden. *Institute of Gas Technology Research Bulletin* No. 33, 1961.

Statistiques Officielles de l'Industrie Gazière en France pour l'Année 1960. Ministère de l'Industrie, Direction du Gaz et de l'Electricité.

Stockage Souterrain des Différents Gaz et Produits Liquides Utilisés par l'Industrie Gazière, by R. Delsol.

Taschenbuch für das Gas-und —Wasserfach, 1961-62, 1 Teil : Jahrbuch ; 2 Teil : Gas.

Translations

The following translations have recently been added to the Library and are available on loan :

Contribution to the Study of the Formation, Elimination and Determination of Traces of Iron Carbonyl and Nickel Carbonyl in Gas from the Underground Reservoir at Beynes, by C. Degent and C. Lebras. (Association Technique de l'Industrie du Gaz en France, Congress, 1961.)

Purification Methods for Various Gases according to Type and Utilization Requirements. (Association Technique de l'Industrie du Gaz en France. Paper IGU/7-61 of 8th International Gas Conference.)

Aslib

Through its memership of Aslib (Association of Special Libraries and Information Bureaux), the Institution is able to make use of the following services :—

Information Service : By its wide knowledge of information sources, Aslib is able to secure information on any subject or to indicate the source from which it may be obtained. If it is contained in a publication, arrangements can usually be made for this to be borrowed.

Commonwealth Index of Unpublished Scientific Translations : Details of translations, completed or in prospect, are notified to Aslib by research organizations, Government Departments and other establishments in the United Kingdom and the principal countries of the Commonwealth. The purpose of the Index is to inform enquirers whether a given item has been translated, and, if so, where and how it may be obtained. Translations received by the Institution are notified to Aslib.

Other services of Aslib include a Register of Specialist Translators and a Register of Specialist Indexers.

In addition, the Institution is represented on the Aslib Fuel and Power Group by the Librarian, and useful contact is maintained with other librarians in this field.

The Secretary of the Institution will be pleased to answer any further enquiries.

General Notices

International Developments in Heat Transfer

The second Conference on International Developments in Heat Transfer, held at Boulder, Colorado, U.S.A., in August, 1961, under the aegis of The Institution of Mechanical Engineers, The American Society of Chemical Engineers and The Institution of Chemical Engineers, is to be repeated in the Lecture Hall, Central Hall, Westminster, London, S.W.1, from 8th to 12th January, 1962.

The basis of the Conference will be papers dealing with the fundamental aspects of heat transfer. There will, however, be papers devoted to technology, and others covering such fields as nuclear energy and aeronautical sciences. Aspects to be covered broadly are radiation, convective heat and mass transfer, condensation and boiling, and conduction and diffusion.

Further particulars and registration form may be obtained from The Secretary, The Institution of Mechanical Engineers, 1, Birdcage Walk, Westminster, London, S.W.1.

Courses in Welding Technology

The School of Welding Technology has arranged the following courses in 1962:—

LONDON:

15th to 18th January:	CO_2 Arc Welding.
29th and 30th January:	Design of Welded Pipework.
12th to 16th February:	Non-destructive Weld Examination.
26th to 28th February:	Welding Metallurgy.
12th to 14th March:	Metal Spraying.
15th and 16th March:	Hard-facing by Welding and Metal Spraying.

The following evening courses have also been arranged: —Design for Welding in Steel Construction (Birmingham—November to March; Bristol and Leeds—January to March) and Detail Design of Welded Structured Steelwork (Manchester—January to March).

Further particulars may be obtained from the School of Welding Technology, 54, Princes Gate, London, S.W.7.

Institution Necktie

The Council of the Institution, in response to a number of requests from members, has arranged for the manufacture of an Institution tie, which will enable members readily to identify other members they meet and with whom they associate both in work and socially.

The design incorporates a motif taken from the Institution's coat of arms and consists of a crown and two crossed torches. The tie is made of silk with the motif on a blue background. (See photograph, p. A26).

In order that the sale of the ties shall be restricted to members of the Institution they will be obtainable only from the headquarters of the Institution or from district sections at 17*s.* 6*d.* each, post free. This price will enable a contribution of 5*s.* for every tie sold to be made to the Benevolent Fund of the Institution.

Orders accompanied by the appropriate remittance, may be addressed to the Secretary, The Institution of Gas Engineers, 17, Grosvenor Crescent, London, S.W.1.

Library

The Library and Reading Room of the Institution are open on weekdays between 9 a.m. and 5.30 p.m. Members may borrow books by post, excepting books of reference and certain others, on application to the Secretary.

Benevolent Fund

Contributions to the Benevolent Fund of the Institution may be paid at any time to the Secretary of the Benevolent Fund, 17, Grosvenor Crescent, London, S.W.1. It is a convenience if annual contributions are made by banker's order, and it is particularly advantageous to the Fund if such contributions are made by Deed of Covenant, whereby the Fund is able to reclaim from the Inland Revenue an appropriate sum in respect of income tax paid by the contributor. Further particulars, with deed forms and banker's orders for completion, are obtainable from the Secretary of the Fund.

Officers of District Sections and Affiliated Associations

Sections

Eastern

Chairman :	K. B. Meggitt, M.I.Gas E., Eastern Gas Board, Gas Works, Ford End Road, Bedford.
Honorary Secretary and Treasurer :	J. Hunter-Rioch, M.B.E., M.I.Mech.E., M.I.Gas E., Eastern Gas Board, 52, Sidney Street, Cambridge.

London and Southern

Chairman :	R. M. Farror, M.I.Gas E., South Eastern Gas Board, Katharine Street, Croydon.
Honorary Secretary and Treasurer :	J. T. Veryard, B.Sc., M.I.Gas E., South Eastern Gas Board, Katharine Street, Croydon.

Manchester and District

Chairman :	N. C. Sturrock, M.I.Gas E., North Eastern Gas Board, 48, James Street, Harrogate.
Honorary Secretary :	J. Castle, B.Sc., M.I.Gas E., North Western Gas Board, Radiant House, Bold Street, Liverpool, 1.
Honorary Treasurer :	A. Harrison, M.I.Gas E., 297, Bramhall Lane South, Bramhall, Cheshire.

Midland

Chairman :	F. Harvey, M.I.Gas E., West Midlands Gas Board, Tame Bridge, West Bromwich Road, Walsall.
Honorary Secretary :	D. K. Green, M.I.Gas E., West Midlands Gas Board, Foleshill Road, Coventry.
Honorary Treasurer :	J. H. Wainwright, M.I.Gas E., West Midlands Gas Board, Powke Lane, Blackheath, Birmingham.

North of England

Chairman :	K. E. Brown, A.M.I.Gas E., Northern Gas Board, Market Street, Newcastle-upon-Tyne.
Honorary Secretary :	W. Cummings, A.M.I.Gas E., Northern Gas Board, 30, Grainger Street, Newcastle-upon-Tyne.
Honorary Treasurer :	A. Field, M.I.Gas E., Northern Gas Board, 19, King Street, Penrith, Cumberland.

Scottish

Chairman :	D. L. Dickson, M.I.Gas E., Scottish Gas Board, Gas Offices, Cumberland Street, Dumfries.
Honorary Secretary and Treasurer :	W. Kirk, M.I.Gas E., Kessington, 29, Union Street, Greenock.

South Western

Chairman :	R. J. Williams, M.I.Gas E., South Western Gas Board, 18, Boscawen Street, Truro, Cornwall.
Honorary Secretary :	F. C. Gay, M.B.E., F.C.I.S., Assoc.I.Gas E., Radiant House, G.P.O. Box No. 141, Bristol, 1.
Honorary Treasurer :	T. A. Nicholson, A.M.I.Gas E., South Western Gas Board, 34, Bridge Street, Taunton.

Wales and Monmouthshire

Chairman :	Elwyn Jones, B.Sc., M.I.Gas E., Bryngas, Penrhys Road, Ystrad, Rhondda.
Honorary Secretary :	J. Powdrill, M.B.E., A.M.I.Gas E., Wales Gas Board, Bute Terrace, Cardiff.
Honorary Treasurer :	O. P. Cronshaw, M.B.E., M.I.Gas E., Wales Gas Board, 93, Wellington Road, Rhyl, N. Wales.

Affiliated Associations

Irish Gas Association

President :	W. A. Campbell, M.I.Gas E., Belfast Corporation Gas Works, Ormeau Road, Belfast.
Director :	N. J. Robertson, Assoc.I.Gas E., Alliance and Dublin Consumers' Gas Company, Dublin, Eire.

British Junior Gas Associations' Joint Council

Chairman :	O. Pincock, B.Sc., South Western Gas Board, 112, Union Street, Torquay.
Honorary Secretary :	L. G. Townsend, M.Sc., M.I.Gas E., North Western Gas Board, Hind Street, Birkenhead, Cheshire.

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(*a*) indicates an abstract.

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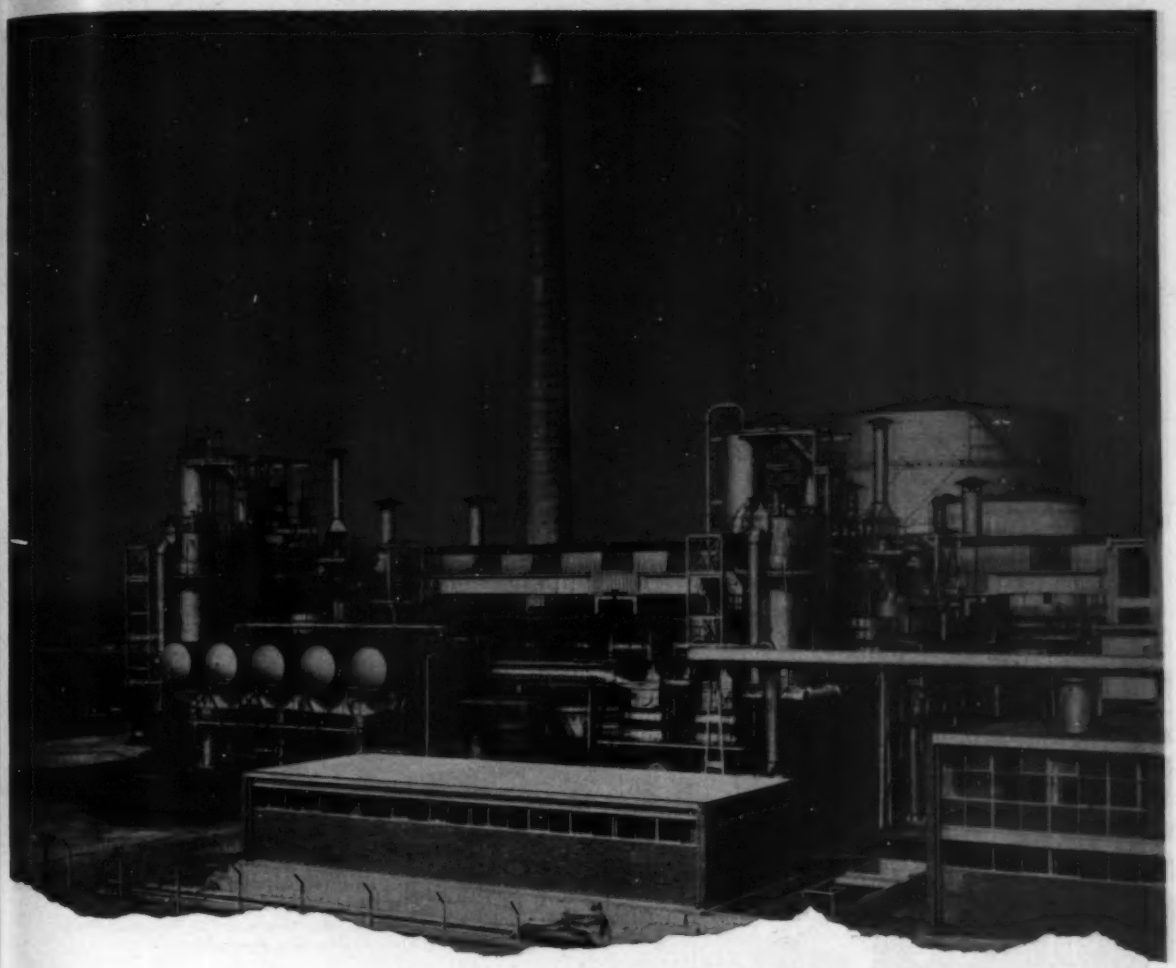

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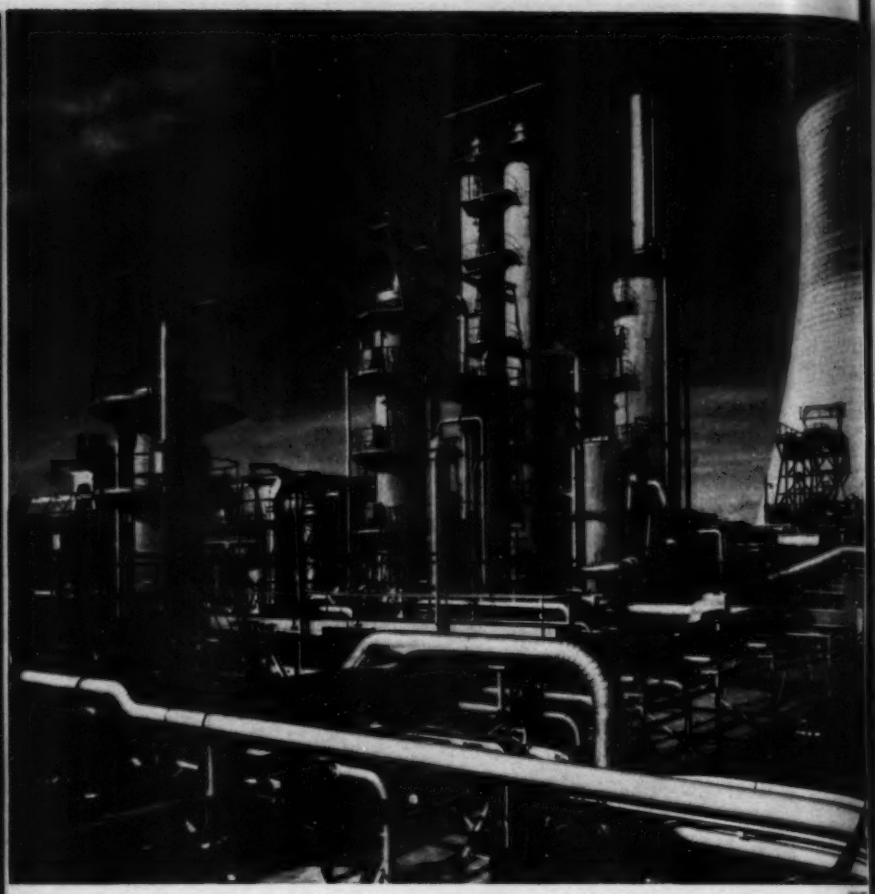

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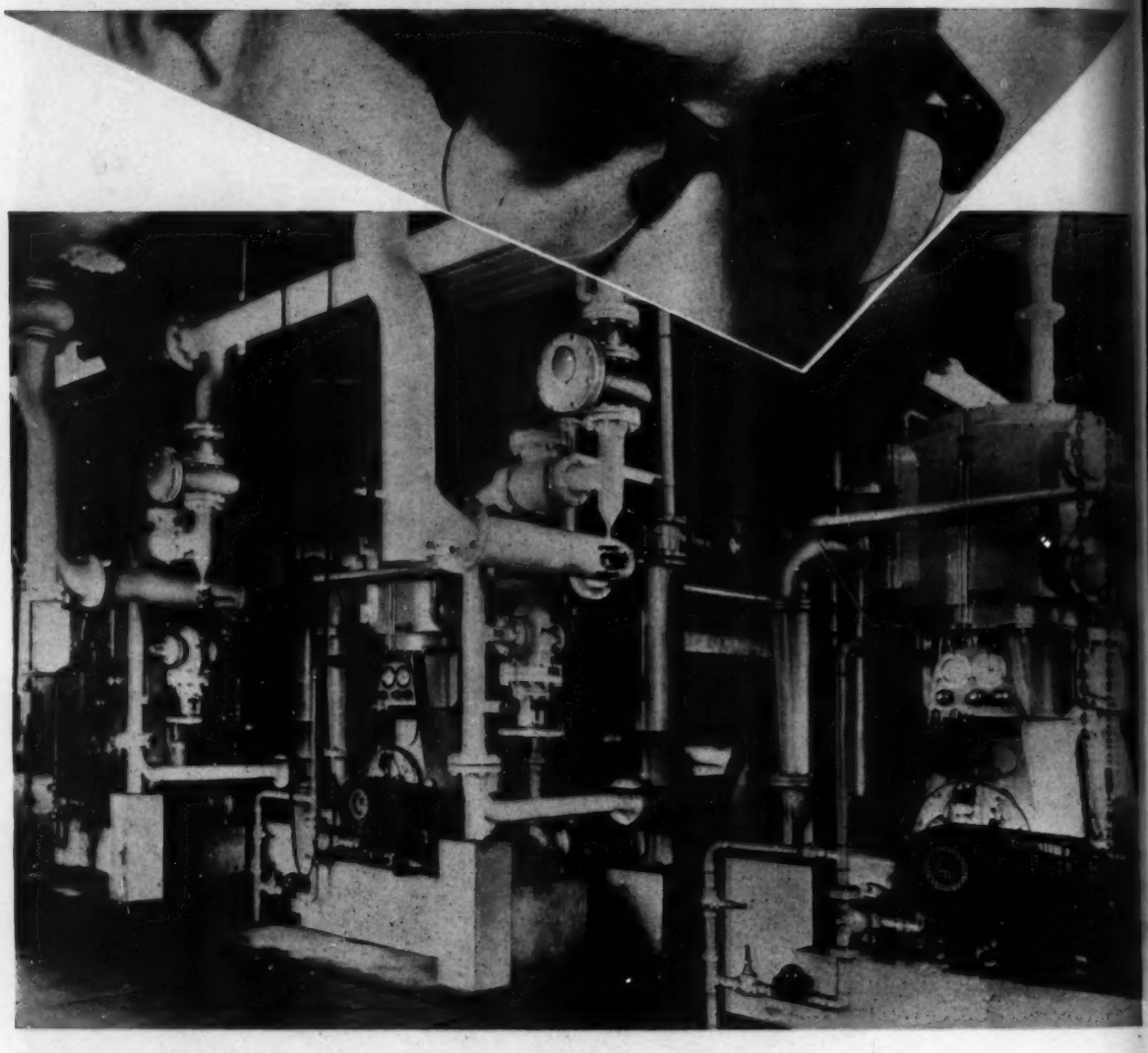
ex COMPRESSORS



Photograph by courtesy of the Wales Gas Board

The photograph shows the installation by The Bryan Donkin Company, Ltd., at the Maelor Works (near Wrexham) of the Wales Gas Board of three Type 300/A 8″ Stroke Two Stage Vertical Reciprocating Compressors. Each compressor is capable of delivering 42,000 c.f.h. at a pressure of 225 p.s.i.g., and driven by a Lister Blackstone EVDF6 flameproof dual-fuel engine. The installation is complete with intercoolers and aftercoolers with valves supplied by The Bryan Donkin Company, Ltd.

The Type 300/A Compressors are employed for the transmission of gas from the Maelor Works through the Mid-Wales Grid as far south as Builth Wells—a distance of approximately 75 miles.

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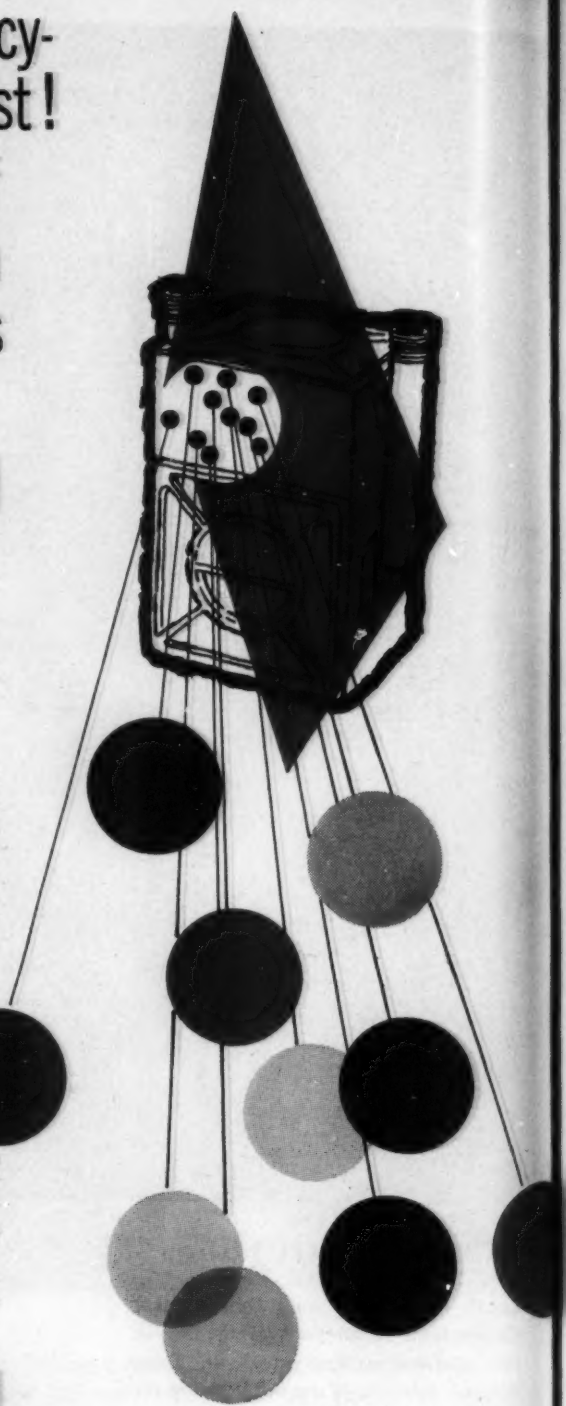
New range of Parkinson Cowan domestic gas meters with **DRY LUBRICATION**



The new Parkinson Cowan dry lubricated domestic gas meters offer under adverse conditions an accuracy of registration previously impossible with greased bearings. These new highly accurate meters are available at the same price as current Parkinson Cowan meters using grease lubrication.

The causes of inaccurate registration

A meter is calibrated when the bearings are new. Stuffing packing is tight and the gear boxes are packed with grease. When the bearings and stuffing boxes are run in and a channel cut in the grease by the revolving gears, the meter will run easier and therefore faster. In time, the grease deteriorates, valves become sticky, bearings wear, corrosion occurs, registration becomes less accurate and revenue drops.

Dry lubrication eliminates these causes

Parkinson Cowan have largely eliminated these causes of inaccurate registration by the introduction of dry lubricated bearings, synthetic rubber seals and stainless steel or hard chrome-plated journals and shafts.

Parkinson Cowan dry lubricated meters have been operated at differential pressures up to nearly four times that experienced under normal service conditions. After large consumptions they have not exhibited the slightest signs of wear.

Further advantages offered by this advanced and proven use of dry lubrication are longer life and greatly reduced maintenance.

[redacted]

MINIMUM MAINTENANCE

Corrosion of bearings caused by interaction between the metal bearings, grease and gas impurities is often the cause of breakdowns and the need for costly maintenance. The effects of corrosion and bearing wear caused by sticky valves are almost completely eliminated from the new Parkinson Cowan range. DRY LUBRICATION NEEDS NO MAINTENANCE.

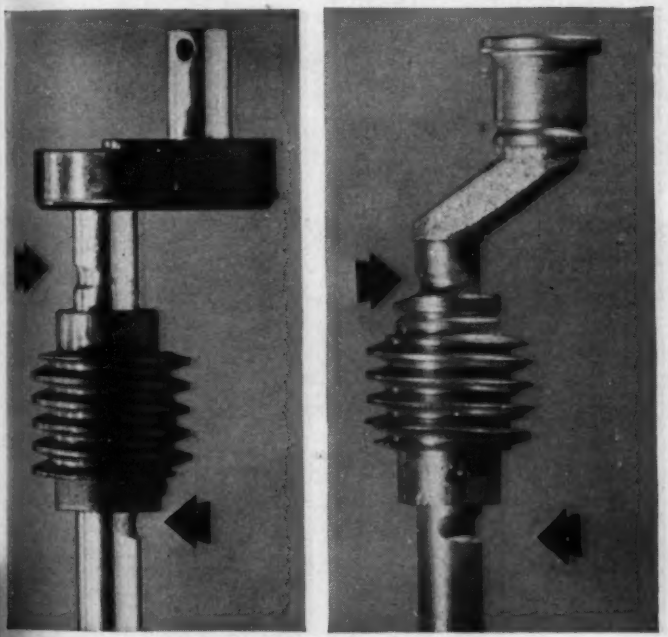
LONGER LIFE *Tests carried out on unpurified gas and under accelerated test conditions to the equivalent of 30 years working life showed the dry lubricated moving parts of a meter to be unaffected by wear or corrosion.*

Compare these crankshafts

These unretouched photographs show crankshafts which were assembled in D.1. meters. Working on air, the meters were operated at 200 cu.ft./hr. with the differential pressure artificially raised by loading the valves and index drive *from 3|10th ins. w.g. to 8|10th ins. w.g.*

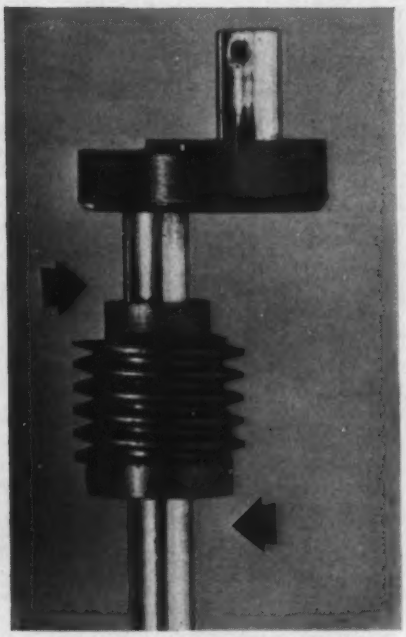
In order to simulate conditions where grease has failed, as it does on certain towns' gas, no grease was used for the test.

40,000 Revolutions *Right.* Bell metal crankshaft, with brass worm and bearing in stainless steel eyelets, after 40,000 revolutions. *Left.* Stainless steel crankshaft with worm and bearing in stainless steel eyelets after 40,000 revolutions.



Note wear after 40,000 revolutions.

500,000 Revolutions Stainless steel crankshaft with Ralsin worm and bearing in Ralsin eyelets after 500,000 revolutions.



Note absence of wear after 500,000 revolutions with dry lubricated bearings.

The Parkinson Cowan range of gas meters—with dry lubrication
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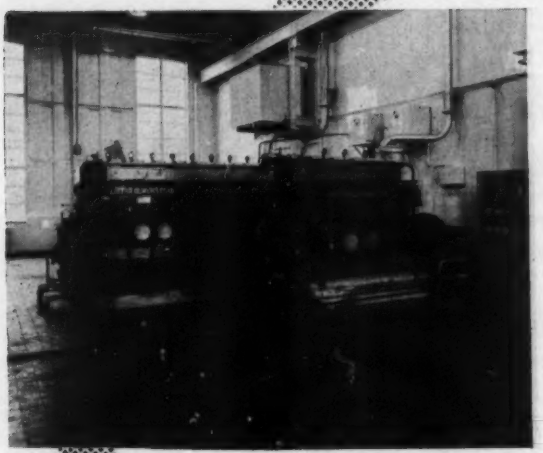
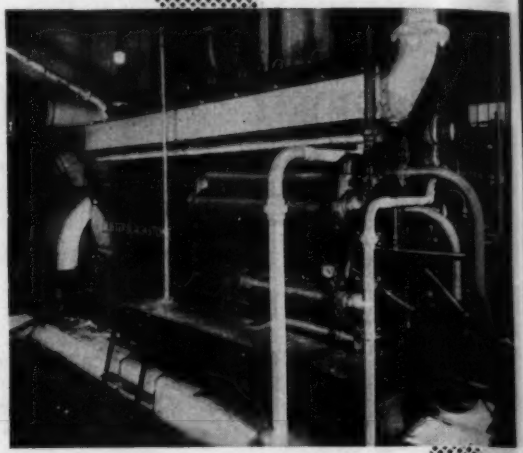
PCGM15

An eight-cylinder and a four-cylinder engine driving Bryan Donkin compressors through a gas-tight wall at the Stapleton Road works of the South-Western Gas Board.



Eight-cylinder dual-fuel engine installed at the Bedwas Works of the Wales Gas Board. Fitted with standard P.A.G.A. equipment.



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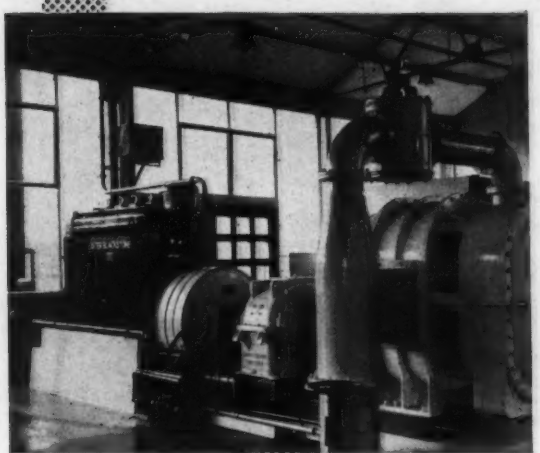
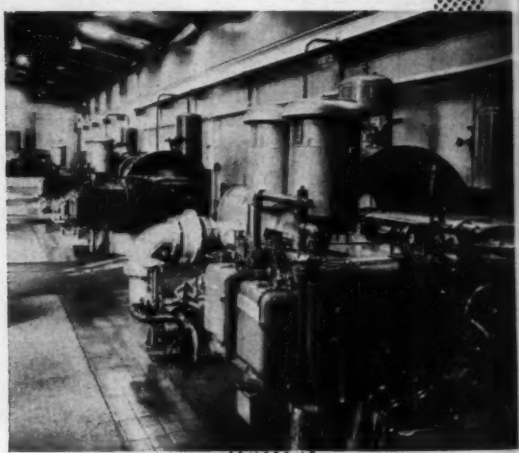


Four-cylinder engine driving a George Waller rotary booster at the James Bridge Boosting Station, Walsall, of the West Midlands Gas Board.



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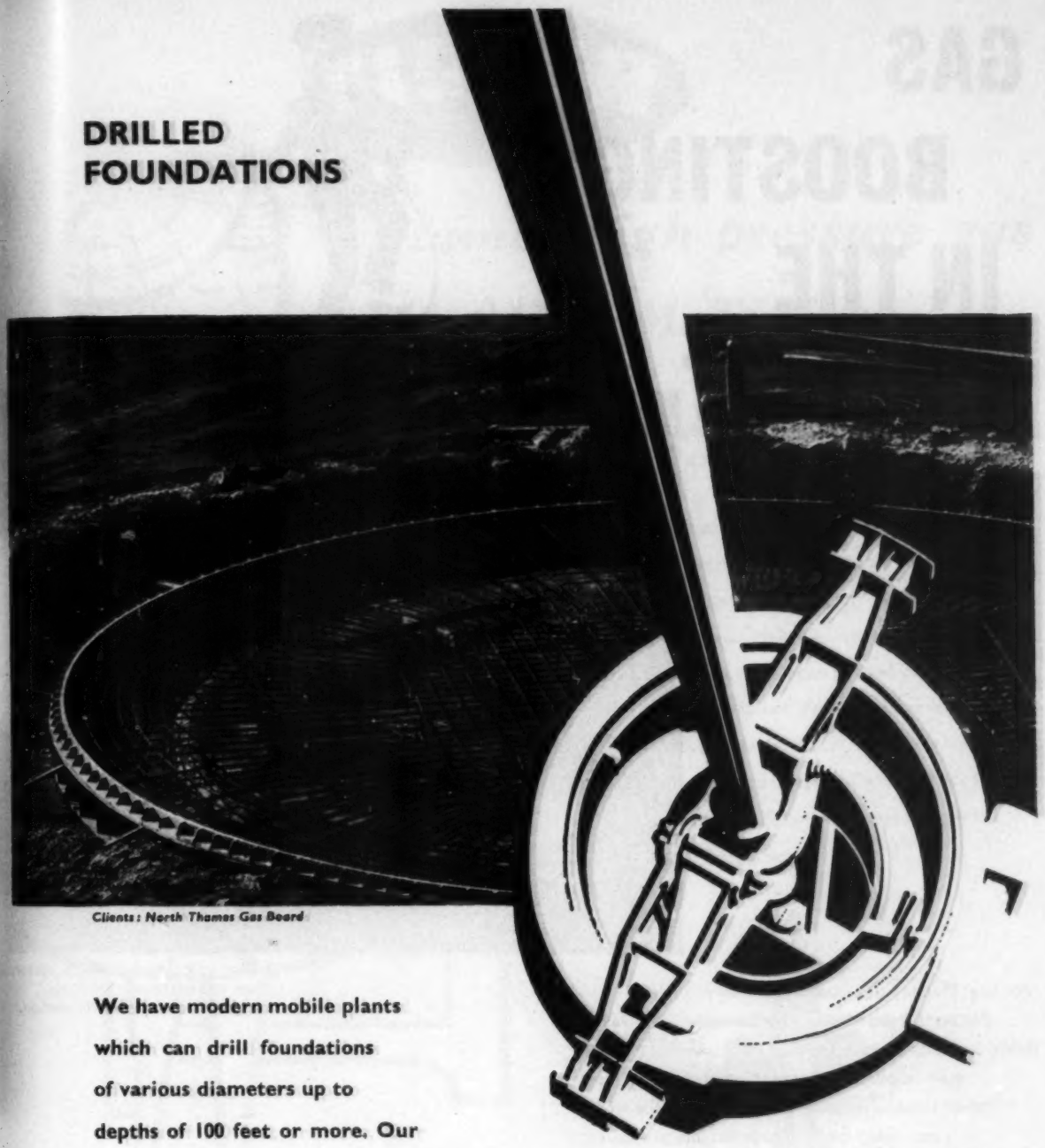

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P.7.5

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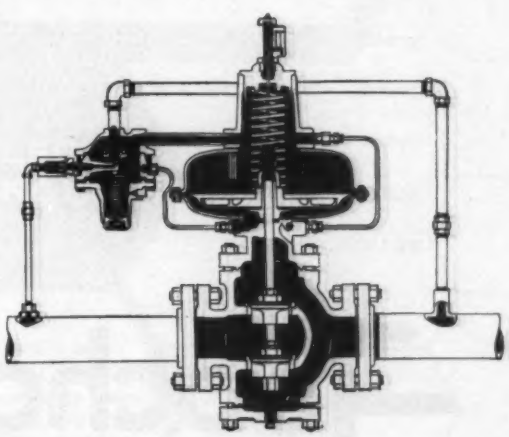

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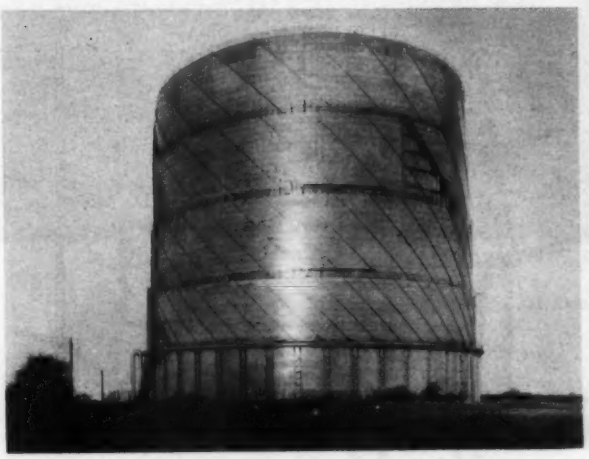

Cowley, Oxford. Photo by Courtesy of Southern Gas Board and C. & W. Walker Ltd.

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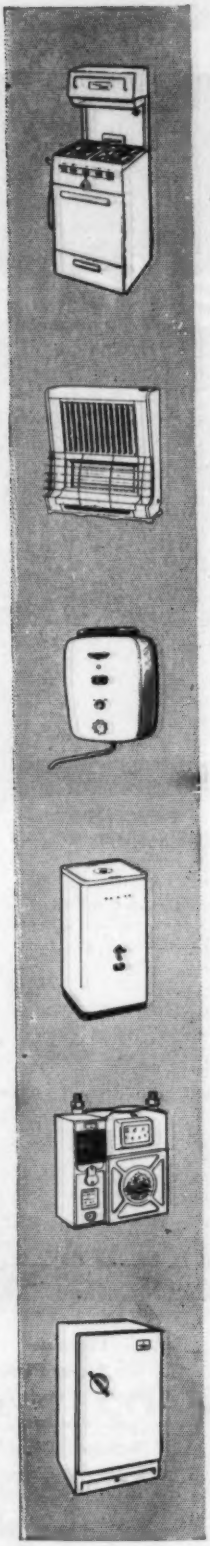

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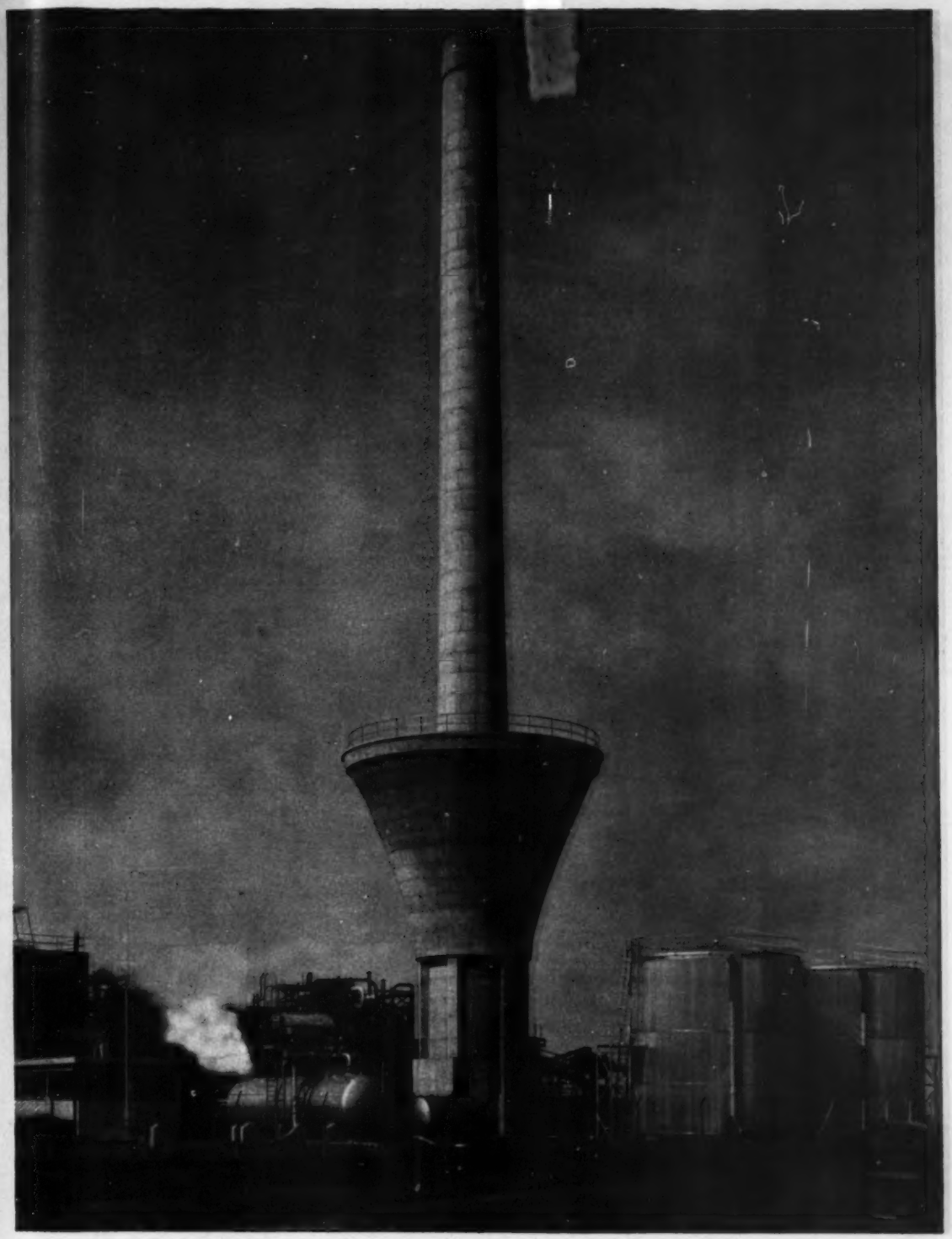

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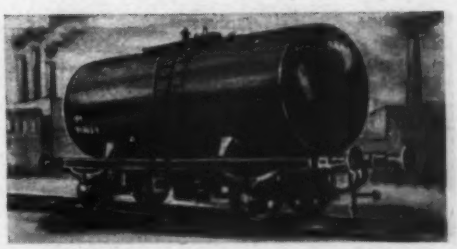

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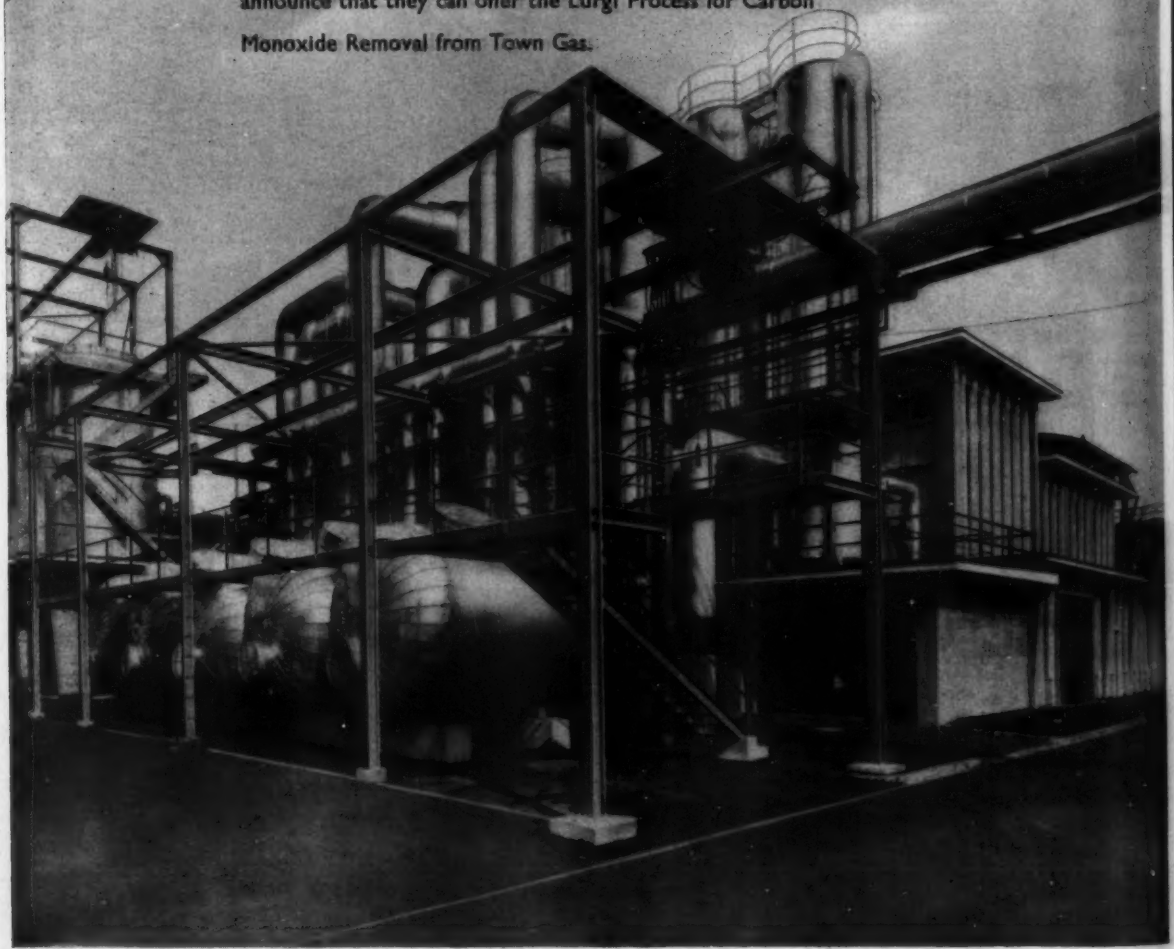

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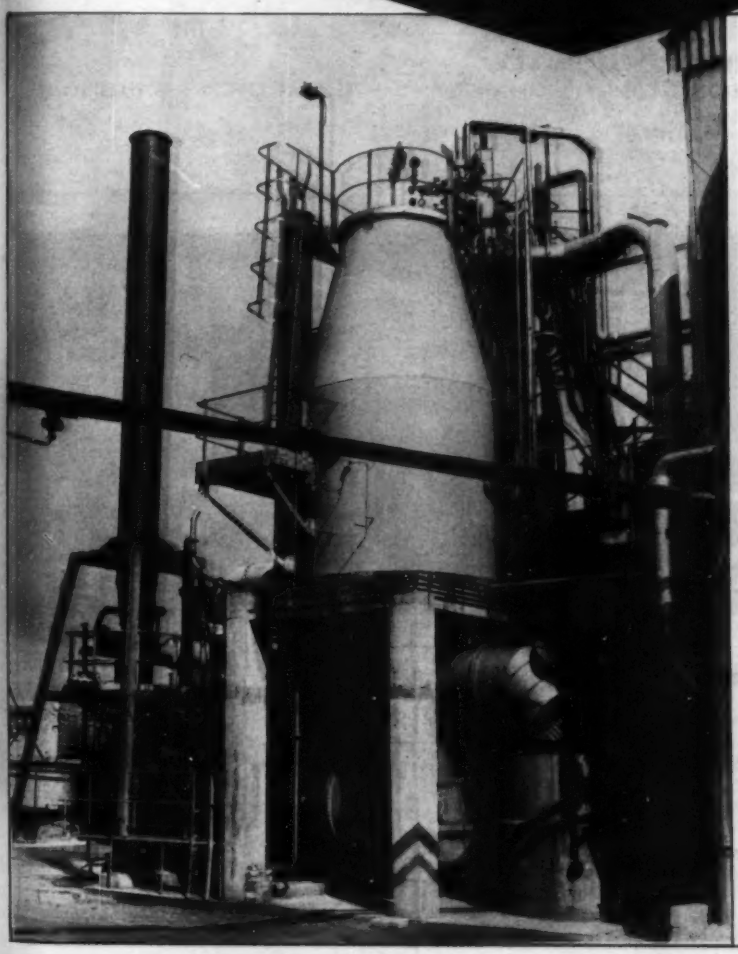

The photograph shows one of the two sets of Onia-Gegi catalytic cracking plants designed and built by Humphreys and Glasgow for the Cork Gas Consumers' Company.

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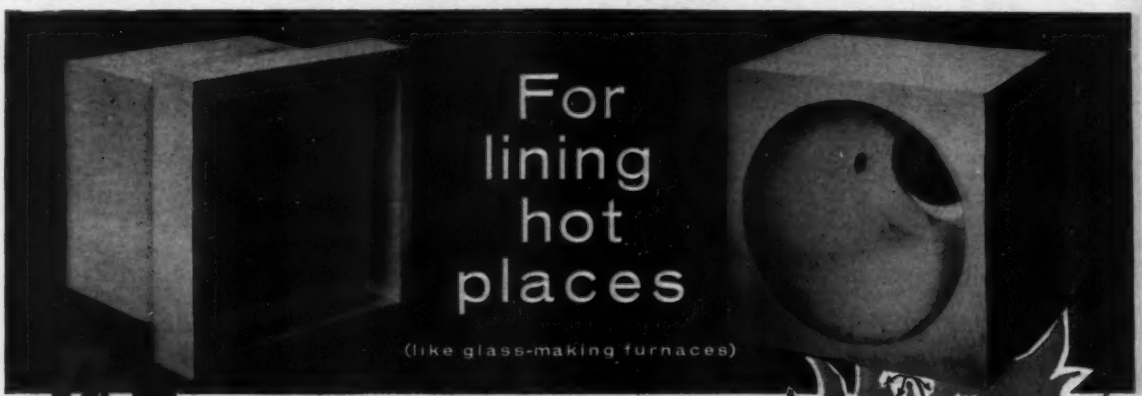

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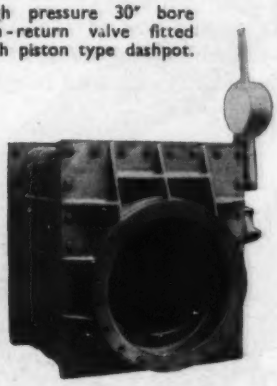

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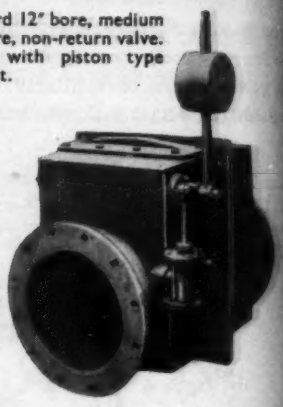

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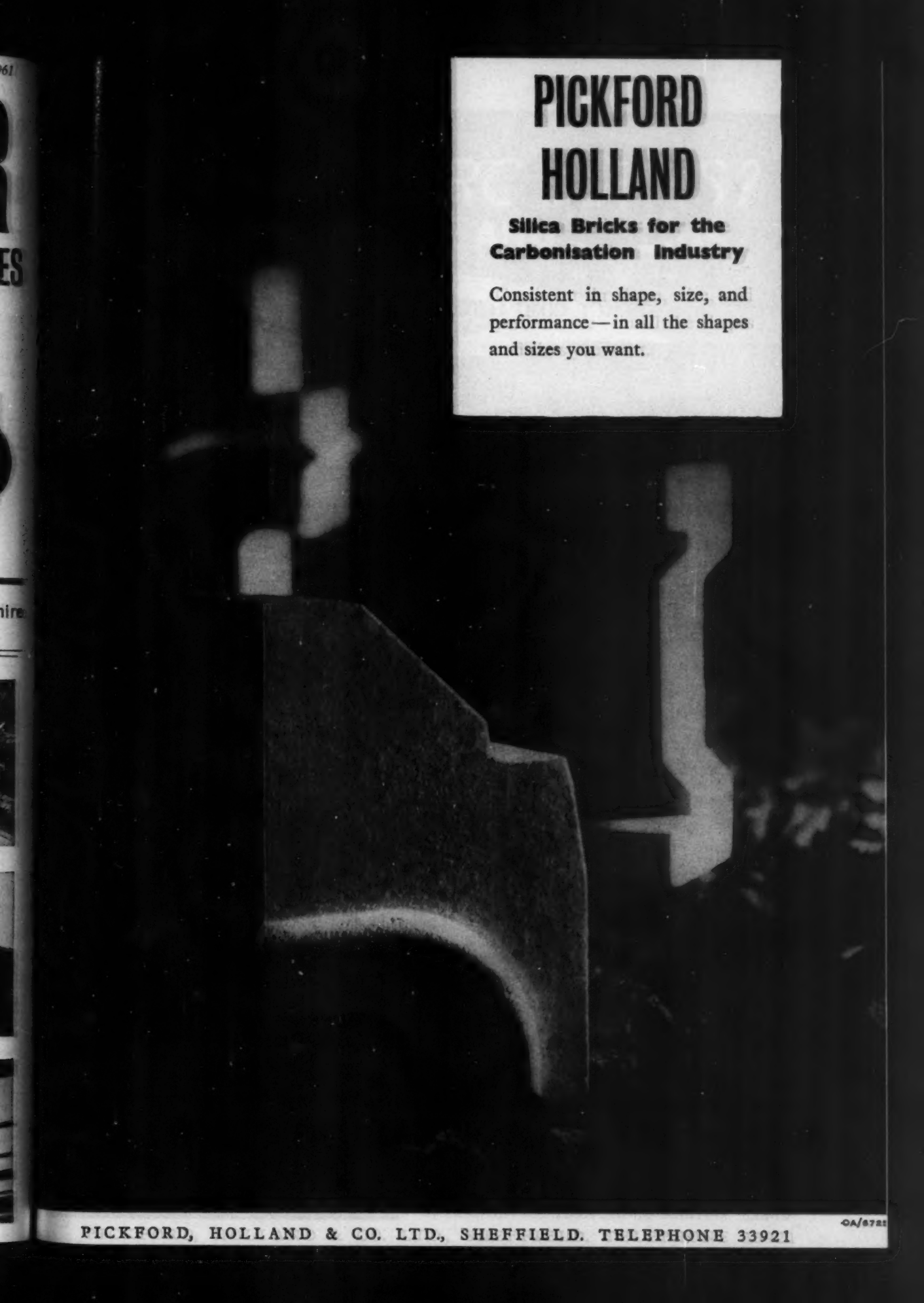
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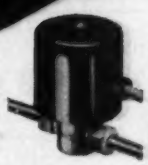

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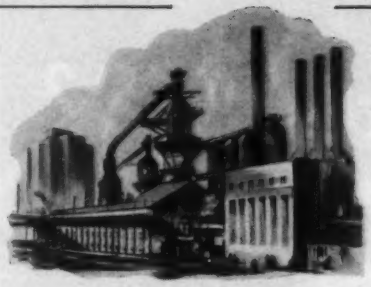

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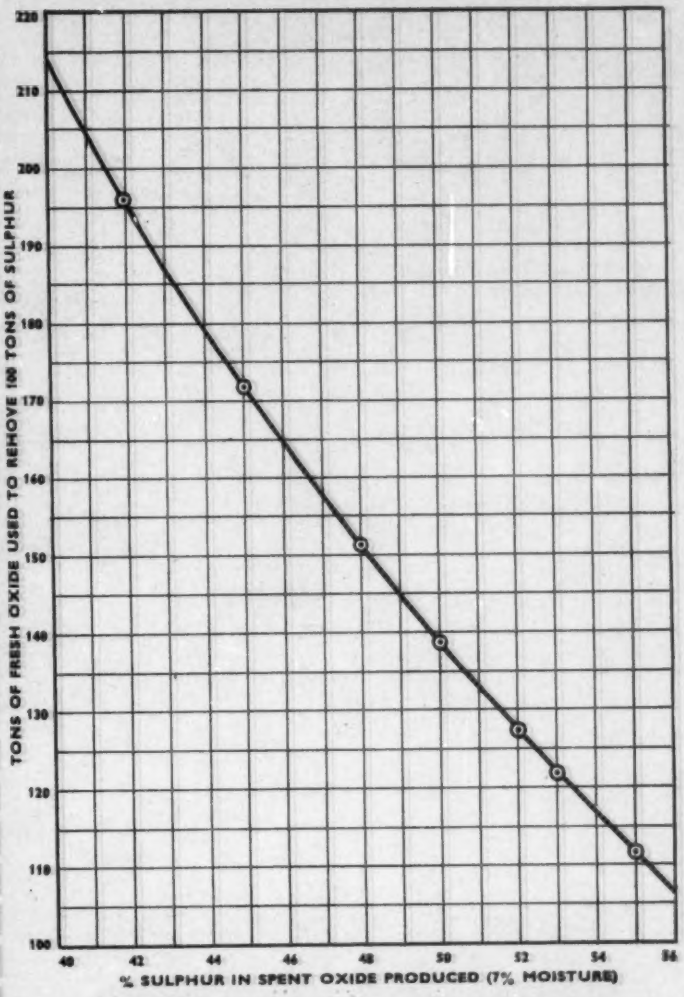

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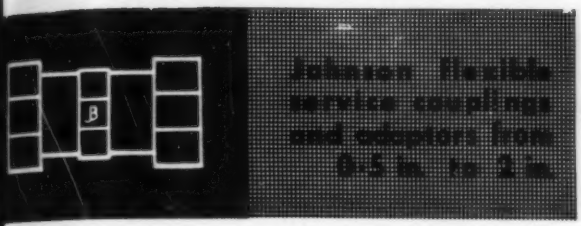

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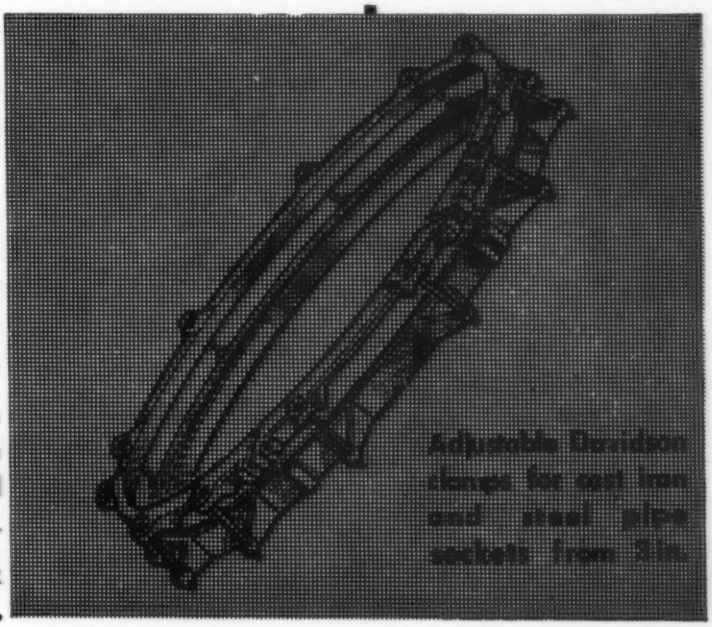

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